

ECONOMIC EVALUATION OF BURNING BITUMINOUS AND WESTERN PRB
COALS FOR POWER PRODUCTION BASED ON SELECT POLLUTION CONTROL
SCENARIOS

BY

JONATHAN PAUL LOFTUS

THESIS

Submitted in partial fulfillment of the requirements
for the degree of Master of Science in Environmental Engineering in Civil Engineering
in the Graduate College of the
University of Illinois at Urbana-Champaign, 2012

Urbana, Illinois

Co-advisors:

Professor Mark J. Rood
Dr. Yongqi Lu

ABSTRACT

The U.S. Environmental Protection Agency (USEPA) finalized its Clean Air Interstate Rule (CAIR) and Clean Air Visibility Rule emission regulations in 2005, to limit emissions of sulfur dioxide, nitrogen oxides and particulate matter to the atmosphere. USEPA replaced CAIR with the Cross State Air Pollution Rule in August 2011. USEPA's Clean Air Mercury Rule regulation was vacated by the D.C. Circuit Court in February of 2008; however, USEPA's "National Emission Standards for Hazardous Air Pollutants from Coal and Oil-Fired Electric Utility Steam Generating Units" proposed in March 2011 further limits emissions of mercury to the atmosphere. As a result of these regulations, this study used the Integrated Environmental Control Model (IECM) to investigate the economic competitiveness of burning select blends of Illinois high-sulfur bituminous and western low-sulfur Powder River Basin (PRB) coals at pulverized coal (PC) power plants while meeting the air emission limitations in these regulations. Since power plants are one of the biggest emission sources of carbon dioxide (CO₂) – a main greenhouse gas – and thus a target for potential CO₂ regulations, 90% CO₂ capture from the flue gas using a monoethanolamine (MEA) absorption process was also evaluated within the IECM study. Most existing CO₂ capture systems utilize absorption-based technology, though it is an energy intensive process. The solid adsorption method has potential to be competitive with the MEA absorption process, regarding energy conversion efficiency for a PC power plant. Therefore, a second part of this study investigated a process using limestone to remove 90% of the CO₂ emissions at a PC power plant. Simulation of the limestone process was performed using ChemCAD, as a "proof-of-concept" study with the goal of estimating the best-case energy use of the process at a PC power plant.

The IECM study showed that the most cost-effective case scenario for PC power plants without CO₂ capture, at the 2007 market coal costs, is an Illinois mine-mouth coal for a 650 MW_e (gross) plant (\$67.0/MWh) and a 70/30 PRB/Illinois coal blend for a 175 MW_e (gross) plant (\$95.0/MWh). The Illinois mine-mouth coal is most cost-effective for the 650 MW_e case due primarily to its lower coal cost compared to the other coal types. The lower coal cost helps compensate for the higher SO₂ control cost for the Illinois mine-

mouth coal compared to the other coal types. The 70/30 PRB/Illinois coal blend replaces the Illinois mine-mouth as most cost-effective when the plant size is reduced to 175 MW_e – even though the Illinois mine-mouth coal still has a lower coal cost compared to the other coal types – for the following reason: the SO₂ and PM control costs for the Illinois mine-mouth case have a higher contribution to plant cost of electricity at the 175 MW_e plant than at the 650 MW_e plant, whereas these control costs for the 70/30 PRB/Illinois coal blend have essentially the same contribution to plant cost at both plant sizes. The most cost-effective case scenario for PC plants seeking to add CO₂ capture, at the 2007 market coal costs, is the 70/30 PRB/Illinois coal blend for both a 650 MW_e (gross) plant (\$133.1/MWh) and a 175 MW_e (gross) plant (\$177.0/MWh). The CO₂ control cost for the 70/30 PRB/Illinois coal blend is higher than the 100% Illinois and Illinois mine-mouth coals, due to the higher CO₂ emissions inherent to burning PRB coal. However, the higher SO₂ control costs for the 100% Illinois and Illinois mine-mouth coals, compared to the 70/30 PRB/Illinois coal blend, counter the lower CO₂ control costs for these coals sufficiently to make the 70/30 PRB/Illinois coal blend most cost-effective for the 650 MW_e and 175 MW_e cases.

The ChemCAD study of the limestone process for CO₂ capture at a 533 MW_e (gross) PC power plant – 498.5 MW_e (net) before installation of CO₂ control – showed that the total auxiliary power use for a best-case scenario of limestone adsorption-desorption is 150 MW_e, compared to 175 MW_e for the MEA process. The power use in the limestone process is attributed primarily to the CO₂ compressor, followed by the air separation unit (ASU) required to calcine the limestone, and then the main feed pump. The power use for CO₂ compression and the main feed pump in the limestone process is higher than the MEA process, and the ASU is required only for the limestone process. However, the power loss due to steam extraction from the power plant steam cycle for MEA regeneration leads to a higher total auxiliary power use for the MEA process. High quality carbonation reaction heat, along with other additional heat from the high-temperature (650 - 950 °C) limestone process, is recovered for producing steam used for additional electricity generation. This additional generation of electricity contributes to higher net generation efficiency at the power plant for the limestone process (30.8%) compared to the MEA process (27.2%).

The IECM portion of this study may be used by managers of coal-fired power plants to assist in determining the most cost-effective approach when burning select coals while simultaneously meeting stack emission regulations. Law- and policy-makers may use results from the IECM study to evaluate mining options to lower free-on-board mine costs, evaluate coal transportation costs, and/or develop policy options concurrent with desired impacts on coal production and sales. Results from the ChemCAD portion of this study can be useful for assessing select CO₂ control technologies regarding impact on energy conversion efficiency.

ACKNOWLEDGEMENTS

Thanks to Dr. Massoud Rostam-Abadi and Prof. Mark J. Rood for providing this research experience for me, and encouraging and supporting me. Thanks to Dr. Yongqi Lu for his steadfast technical guidance, without which the completion of this research would not have been possible. Thanks also to my family and friends for their encouragement and support, especially to my mother for always being there for me.

TABLE OF CONTENTS

	PAGE No.
LIST OF TABLES.....	viii
LIST OF FIGURES	ix
1 INTRODUCTION	1
1.1 Coal Consumption in the World and United States.....	1
1.2 Emission Regulations in the U.S. and Importance of Multi-pollutant Control Decisions.....	3
1.3 Bituminous and Sub-bituminous Competitive Edge	4
1.4 Potential CO ₂ Control	5
1.4.1 CO ₂ Emissions and Potential Regulation.....	5
1.4.2 CO ₂ Control Methods	7
1.5 Coal Costs	13
1.6 Introduction to Integrated Environmental Control Model.....	13
1.7 Objectives	16
1.8 Research Justification	16
2 METHODS	18
2.1 IECM Evaluation of COE.....	18
2.1.1 Coal Specifications	18
2.1.2 Environmental Control Configurations.....	19
2.1.3 Scenarios.....	19
2.2. Assessment of Limestone Carbonation-Calcination Process.....	23
2.2.1 Description of Reference Plants.....	23
2.2.2. Limestone Carbonation-Calcination Process.....	26
2.2.3. Simulation of LSCC Process	36
3 RESULTS AND DISCUSSION.....	39
3.1 Scenario 1 - SO ₂ , NO _x , PM and Mercury Control with no CO ₂ Capture	39
3.2 Scenario 2 - SO ₂ , NO _x , PM and Mercury Control and 90% CO ₂ Capture with MEA	47
3.3 Assessment of Limestone Carbonation-Calcination Process.....	55
4 SUMMARY OF CONCLUSIONS AND RECOMMENDATIONS AND FUTURE WORK	61
4.1 IECM Evaluations.....	61
4.2 LSCC Process Simulation.....	63
REFERENCES	66
APPENDICES	72
Appendix A General Items for LSCC Process	72
Appendix A1 Equilibrium Curve for CaO with CO ₂	72
Appendix A2 Equilibrium Curve for MgO with CO ₂	73
Appendix A3 Equilibrium Curve for Na ₂ CO ₃ with CO ₂	74
Appendix A4 Steam Utilization Curve	75
Appendix B Assumptions for ChemCAD Simulations of Base Plant and Base Plant with MEA Process	76
Appendix B1 Operating Conditions of Base Plant	76
Appendix B2 Operating Conditions for MEA Process.....	77

Appendix B3 Energy Use for Circulating Water Pumps and Cooling Tower Fans .	78
Appendix C Assumptions for ChemCAD Simulation of LSCC Process	79
Appendix C1 Boiler	79
Appendix C2 Operating Conditions for LSCC Process.....	81
Appendix C3 Carbonator	83
Appendix C4 Calciner.....	85
Appendix D Impacts of CO ₂ Capture on PC Power Plants (650 MW _e)	87
Appendix E Power Plant with 533 MW _e (Gross) Output Installed with the LSCC Process – Simulation Results of Mass and Energy Balance Calculations from ChemCAD.....	93

LIST OF TABLES

Table 1 Pollutant emission limits or practices	4
Table 2 Current and future technologies for CO ₂ capture	9
Table 3 Control technologies used in IECM.....	15
Table 4 Coal specifications for the scenarios tested	18
Table 5 Model parameters and values for scenarios	20
Table 6 Qualitative comparison of LSCC and MEA processes.....	35
Table 7 Main operating and performance parameters for MEA and LSCC processes.....	37
Table 8 Breakeven delivered Illinois coal costs against a fixed PRB cost of \$27.5/short ton (\$30.3/metric ton) (PC-tangential fired boiler).....	42
Table 9 Breakeven delivered Illinois coal costs against a fixed PRB cost of \$27.5/short ton (\$30.3/metric ton) (PC-tangential fired boiler with CC)	50
Table 10 Auxiliary power use and overall process performance for select cases	56
Table 11 Operating conditions of base plant	76
Table 12 Simulation Parameters for MEA Process	77
Table 13 Properties of Illinois coal ¹⁴	79
Table 14 Simulation parameters for LSCC process.....	81
Table 15 Impacts of CO ₂ capture on PC power plant (650 MW _e gross)	87
Table 16 Impacts of CO ₂ capture on PC power plant (175 MW _e gross)	90
Table 17 Boiler stream conditions	94
Table 18 Stream conditions for steam cycle	96
Table 19 Stream conditions for LSCC process.....	98
Table 20 Stream conditions for CO ₂ compression.....	100
Table 21 Stream conditions for compression of different gas streams.....	102

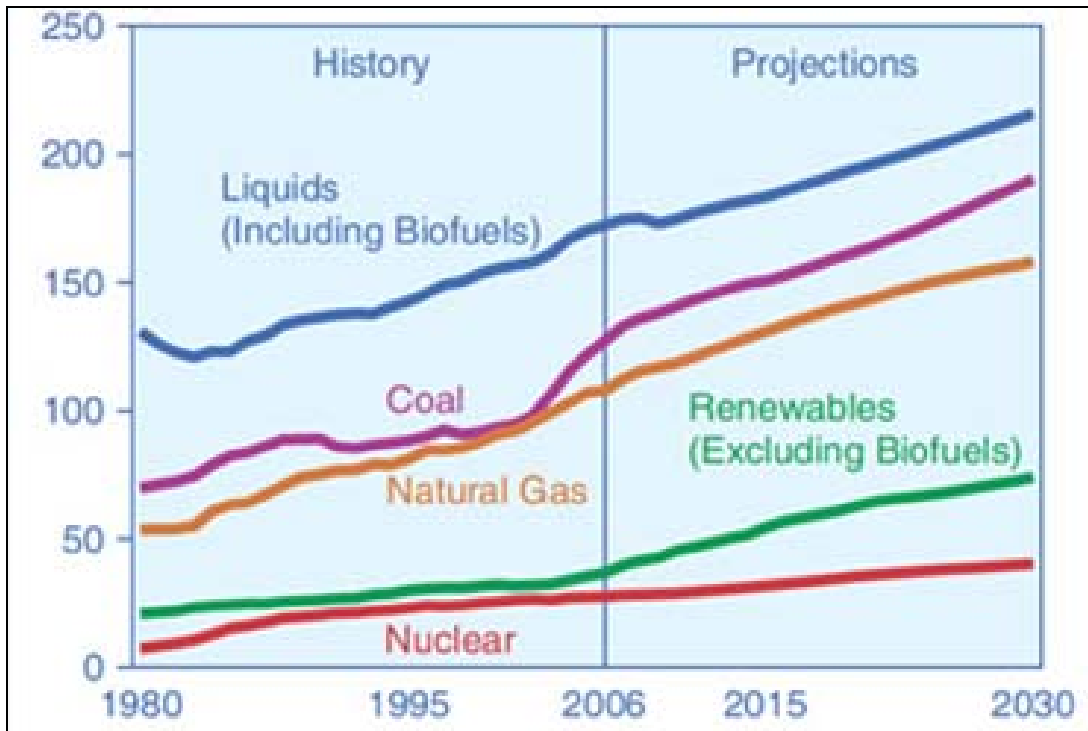
LIST OF FIGURES

Figure 1 World marketed energy use by fuel type on vertical axis (quadrillion British thermal units) vs. year on horizontal axis	1
Figure 2 Primary energy use in U.S. by fuel type on vertical axis (quadrillion British thermal units) vs. year on horizontal axis	2
Figure 3 World energy-related CO ₂ emissions by fuel type (billion metric tons) on vertical axis vs. year on horizontal axis	5
Figure 4 CO ₂ emissions in the U.S. in 2006 and 2030 (million metric tons) on vertical axis vs. sector and fuel on horizontal axis	6
Figure 5 Three main processes that may be employed to produce power from coal and then capture the CO ₂ emissions	8
Figure 6 Environmental control configurations for different coals	19
Figure 7 Schematic of base power plant with no CO ₂ capture	24
Figure 8 Schematic of reference power plant with MEA process.	25
Figure 9 Schematic of overall power plant with LSCC process	28
Figure 10 Schematic of LSCC process	29
Figure 11 COE versus delivered Illinois coal cost for 650 MW _e (gross) plant	40
Figure 12 COE versus delivered Illinois coal cost for 175 MW _e (gross) plant	41
Figure 13 Breakdown costs for 650 MW _e (gross) PC plant	44
Figure 14 Breakdown costs for 175 MW _e (gross) PC plant	46
Figure 15 COE versus delivered Illinois coal cost for 650 MW _e (gross) PC-CC plant....	48
Figure 16 COE versus delivered Illinois coal cost for 175 MW _e (gross) PC-CC plant....	49
Figure 17 Breakdown costs for 650 MW _e (gross) PC-CC plant.....	52
Figure 18 Breakdown costs for 175 MW _e (gross) PC-CC plant.....	54
Figure 19 Equilibrium curve for CaO with CO ₂	72
Figure 20 Equilibrium curve for MgO with CO ₂	73
Figure 21 Equilibrium curve for Na ₂ CO ₃ with CO ₂	74
Figure 22 Steam utilization curve	75
Figure 23 Boiler	93
Figure 24 Steam cycle.....	95
Figure 25 LSCC process	97
Figure 26 CO ₂ compression	99
Figure 27 Compression of different gas streams	101

1 INTRODUCTION

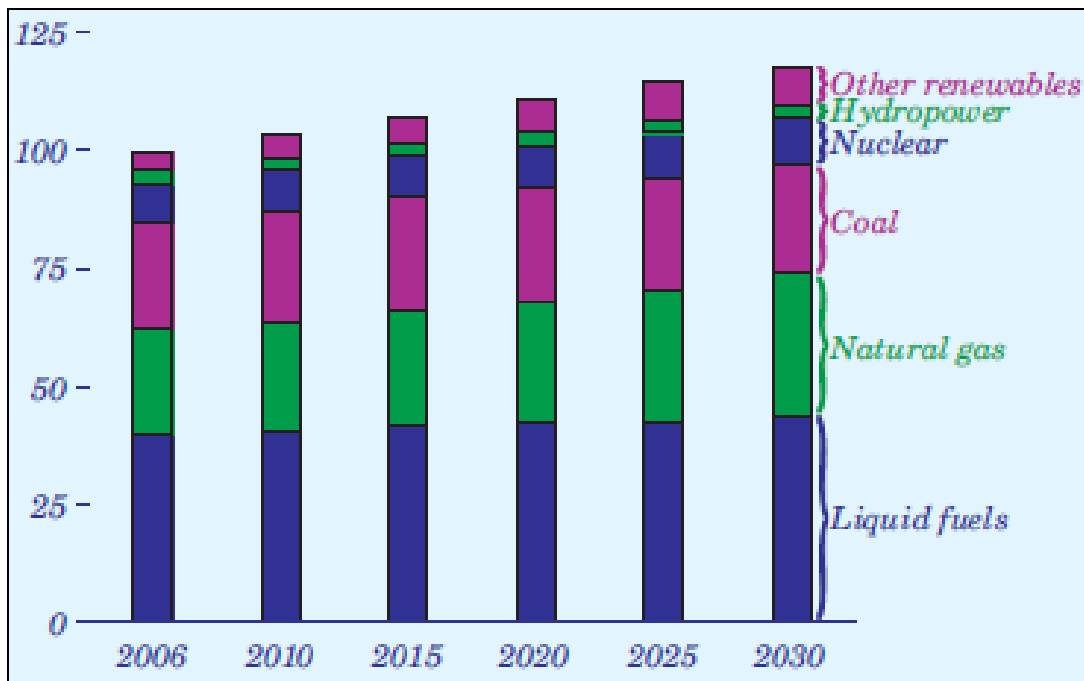
1.1 Coal Consumption in the World and United States

Coal is second, only to liquids (including biofuels), as a primary source of energy in meeting daily needs throughout the world, such as electricity, heat, transportation, and industrial operations. Figures 1 and 2 show coal contribution to energy use worldwide from 1980 to 2030 and in the U.S. from 2006 to 2030, respectively.



Note: Figure accessed from U.S. Energy Information Administration¹

Figure 1 World marketed energy use by fuel type on vertical axis (quadrillion British thermal units) vs. year on horizontal axis



Note: Figure accessed from U.S. Energy Information Administration²

Figure 2 Primary energy use in U.S. by fuel type on vertical axis (quadrillion British thermal units) vs. year on horizontal axis

There were a total of 472 quadrillion British thermal units (Btu) – or 498 quadrillion kilojoules (kJ) – of fuel consumed worldwide in 2006.¹ Liquid fuels, which hold the majority of energy use worldwide and in the U.S., are consumed primarily in the transportation sector, and to a lesser extent in industrial applications. In 2006, natural gas contributed about 15% less to energy use than coal worldwide, and about the same energy use in the U.S.^{1,2} Natural gas is used primarily in the electric power and industrial sectors, and to a lesser extent in the commercial and residential sectors. There were 130 quadrillion Btu (137 quadrillion kJ) of coal (6.1 gigatons short, or 5.5 gigatons metric) consumed worldwide in 2006, with the U.S. consuming 22.5 quadrillion Btu (23.8 quadrillion kJ) of coal (1.1 gigatons short, or 1.0 gigatons metric), which is 17% of the worldwide coal use.^{2,3} In 2007, 92% of the coal mined in the U.S. was used for electric power generation within the U.S. : 44.5% of that coal was bituminous coal and 40% was sub-bituminous coal.⁴

1.2 Emission Regulations in the U.S. and Importance of Multi-pollutant Control Decisions

The U.S. Environmental Protection Agency (USEPA) finalized its Clean Air Interstate Rule (CAIR) and Clean Air Visibility Rule (CAVR) emission regulations in 2005. These regulations were created to limit the annual emissions of sulfur dioxide (SO₂), nitrogen oxides (NO_x) and particulate matter (PM) to the atmosphere. In December 2008, the D.C. Circuit Court remanded CAIR without vacatur, which left the CAIR rules in place until USEPA finalized a replacement called the Cross State Air Pollution Rule (CSAPR) in August 2011. USEPA's Clean Air Mercury Rule (CAMR) regulation was vacated by the D.C. Circuit Court in February of 2008. However, USEPA's "National Emission Standards for Hazardous Air Pollutants from Coal and Oil-Fired Electric Utility Steam Generating Units" proposed in March 2011 further limits emissions of mercury to the atmosphere.

The work presented here focused on the environmental regulations to be complied with by 2013 and 2015 at coal-fired power plants, using Illinois regulations as a reference (Table 1).^{5,6}

Table 1 Pollutant emission limits or practices

Pollutant component	Illinois Environmental Protection Agency (IEPA) ⁵		USEPA ⁶
	2013	2015	New Source Performance Standards (NSPS)
SO ₂	0.33 lb/mmBtu (1.4×10^{-4} kg/MJ) or 44% reduction from base rate of emissions	0.25 lb/mmBtu (1.1×10^{-4} kg/MJ) or 55% reduction from base rate of emissions	2.0 lb/MWh (0.91 kg/MWh) gross, or 0.25 lb/mmBtu (1.1×10^{-4} kg/MJ) ^a
NO _x	0.11 lb/mmBtu (4.7×10^{-5} kg/MJ) or 80% reduction from base rate of emissions	Same as 2013	1.0 lb/MWh (0.45 kg/MWh) gross, or 0.126 lb/mmBtu (5.4×10^{-5} kg/MJ) ^a
PM	0.03 lb/mmBtu (1.3×10^{-5} kg/MJ)	0.03 lb/mmBtu (1.3×10^{-5} kg/MJ)	6.4 mg/J, or 0.015 lb/mmBtu
Mercury	8×10^{-6} lb/MWh (3.6×10^{-6} kg/MWh) gross or 90% reduction or injection of halogenated activated carbon ^{a,b}	Same as 2013	2.1×10^{-5} lb/MWh (9.5×10^{-6} kg/MWh) gross ^a
^a “Gross” indicates <i>gross electrical output</i> , which means the total electrical output from an electric generating unit (EGU) before making any deductions for energy output used in any way related to the production of energy (for an EGU generating only electricity, the gross electrical output is the output from the turbine/generator set) ^b For Powder River Basin (PRB) sub-bituminous coal, halogenated activated carbon injection (ACI) rate for mercury control is 5 lb/MMacf, or 8.0 kg/10 ⁶ actual cubic meter (acm), and for bituminous coal the ACI rate is 10 lb/MMacf (15.9 kg/MMacm)			

Even with the D.C. Circuit Court vacatur of CAMR in 2008, the mercury emission standard in Illinois began July 1st, 2009 – 17 other states have also implemented mercury emission rulings. Since the air quality control technology for one particular pollutant can affect other pollutants as well, power plants in the U.S. will likely choose to make decisions soon about their multi-pollutant control scheme.

1.3 Bituminous and Sub-bituminous Competitive Edge

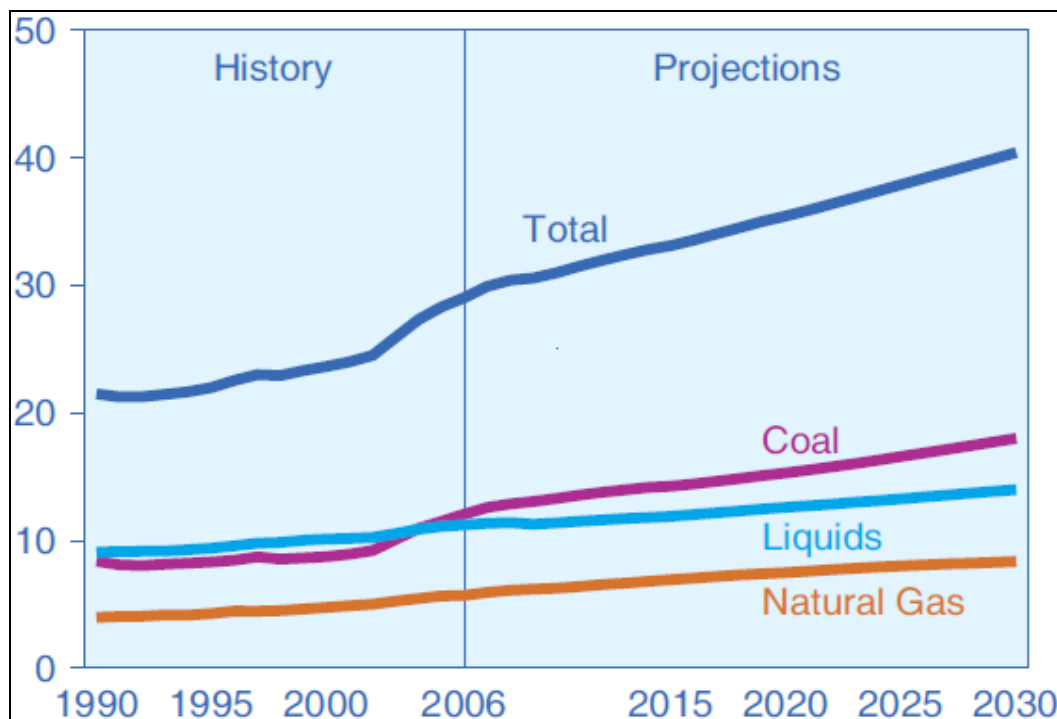
Due to these current and new regulations, the competitive edge between western Powder River Basin (PRB) low sulfur sub-bituminous coal and high sulfur bituminous coal may change. Two major reasons for this possible change are: 1) new SO₂ emission control imposed to the low sulfur coal adds the additional cost of flue gas desulfurization (FGD)

to power plants burning PRB coal; and 2) the inherent mercury oxidation capacity of bituminous coal makes mercury capture from the flue gas more feasible, thus making bituminous coal and PRB/bituminous coal blends possibly more competitive with PRB coal. Also, tighter limits on NO_x and PM emissions may have a significant influence to the competitiveness between these coals under multi-pollutant control schemes.

1.4 Potential CO₂ Control

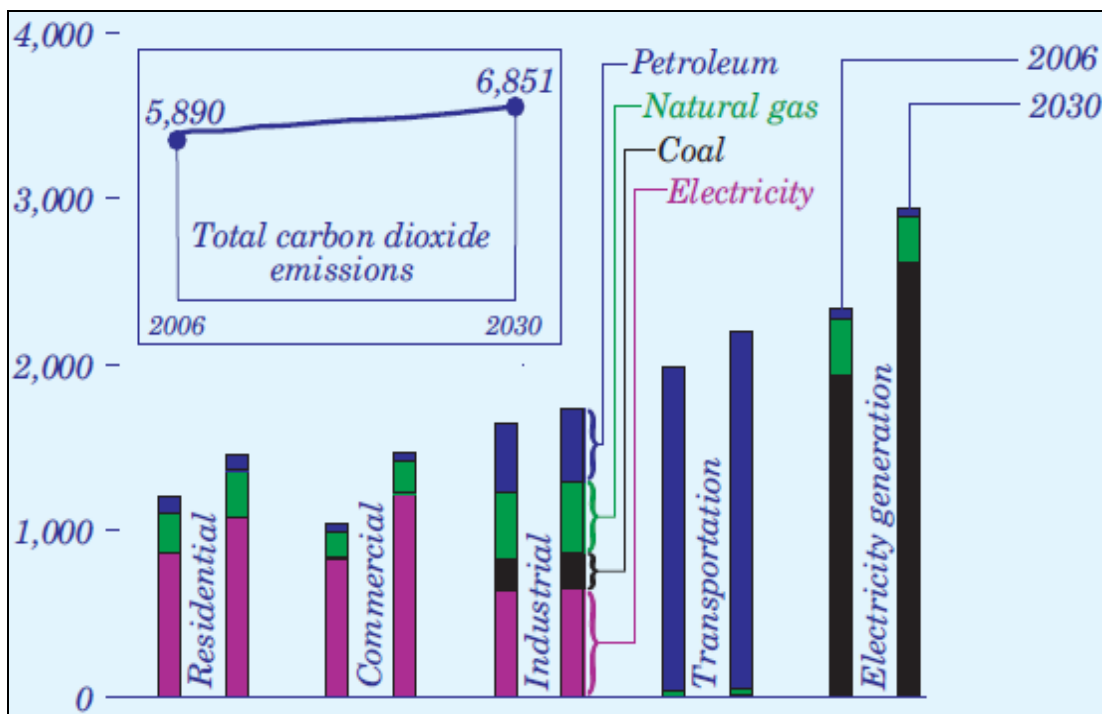
1.4.1 CO₂ Emissions and Potential Regulation

Heat-trapping gases, or greenhouse gases (GHGs) – mainly carbon dioxide (CO₂), methane (CH₄) and nitrous oxide (N₂O) – in the atmosphere due to anthropogenic activities contribute to global warming potential. The atmospheric concentration of CO₂ was at 392 parts per million (ppm) (mol fraction) in July 2011, up from 316 ppm (mol fraction) in 1959, with CH₄ and N₂O showing similar trends in increasing concentrations.⁷ Figures 3 and 4 show CO₂ emissions from consumption of different fuel types worldwide and in the U.S.



Note: Figure accessed from U.S. Energy Information Administration⁸

Figure 3 World energy-related CO₂ emissions by fuel type (billion metric tons) on vertical axis vs. year on horizontal axis



Note: Figure accessed from U.S. Energy Information Administration²

Figure 4 CO₂ emissions in the U.S. in 2006 and 2030 (million metric tons) on vertical axis vs. sector and fuel on horizontal axis

The energy sector – which includes production, transformation, handling, and consumption of energy commodities – is the largest contributor to CO₂ emissions. Energy-related CO₂ emissions in the U.S. was 6.02 billion metric tons (6.64 billion short tons) in 2004, and made up 22% of global energy-related CO₂ emissions.⁹ In 2006, the electricity generation sector contributed 2.33 billion metric tons (2.57 billion short tons), or 39%, of CO₂ emissions in the U.S.¹ Flue gas from coal-fired power plants contains 12-15% CO₂ by volume, and these plants are the main type of electricity generator in the U.S. – 1.75 billion metric tons (1.93 billion short tons) CO₂, or 81% of the electricity generation sector.¹⁰

These large emissions from stationary sources make coal-fired power plants a prime candidate for control if CO₂ is regulated in the future. USEPA finalized its GHG Tailoring Rule in May 2010, under which facilities responsible for nearly 70 % of the national GHG emissions will be subject to permitting requirements. Although these permitting requirements do not necessarily establish GHG control measures for existing

sources, national GHG emission reduction goals may require significant GHG emission reductions from existing coal-fired power plants. One such goal is the American Clean Energy and Security Act of 2009, which is designed to reduce GHG emissions from select sources – electric generators and industrial sources, petroleum and fluorinated gas producers and importers, and local natural gas distribution companies – 17% below 2005 levels by 2020 and 83% below 2005 levels by 2050.¹¹ A regulation to limit CO₂ emissions is expected to favor the use of bituminous coal: the addition of CO₂ capture from the flue gas makes the bituminous coal more competitive with PRB coal, because bituminous coal produces CO₂ emissions (mass CO₂ produced per unit energy produced) at a lower flue gas flow rate than PRB coal.

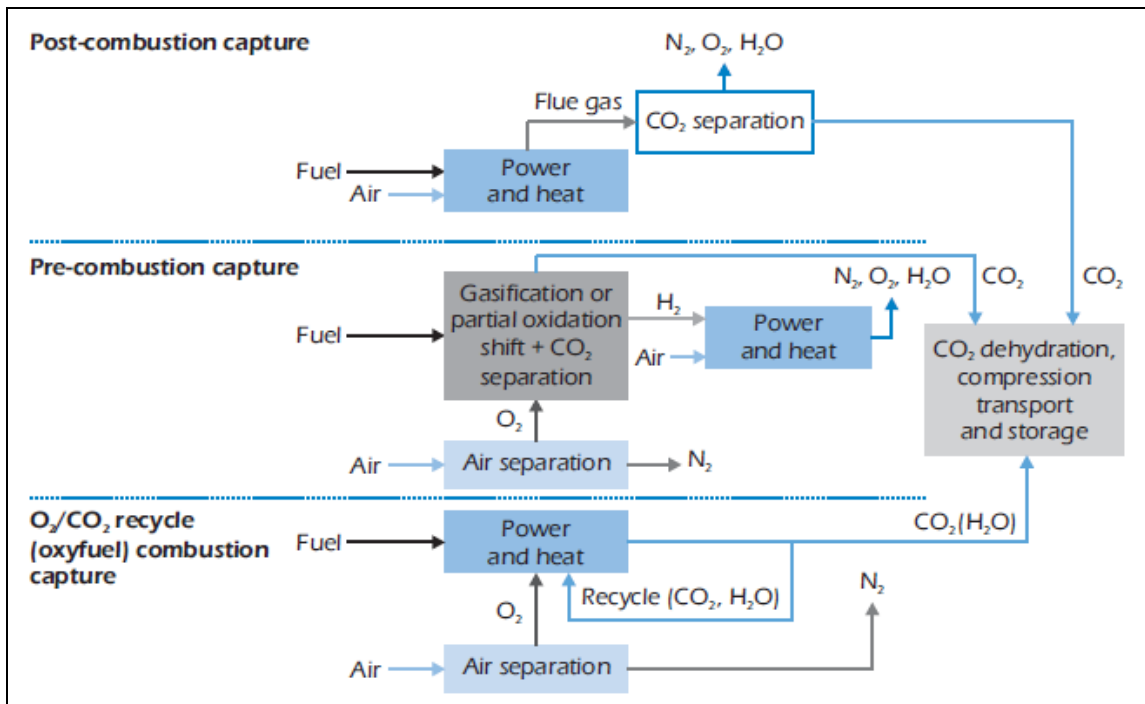
1.4.2 CO₂ Control Methods

Sources of CO₂ emissions include transportation, industrial, commercial, residential, and electricity generation. Among stationary sources, power plants emit the majority of CO₂ (Figure 4). Demand Side Management (DSM) is a primary consideration when evaluating emission reduction methods. DSM refers to all changes that originate from the demand side of the market in order to achieve large scale energy efficiency improvements by deployment of improved technologies. The International Energy Agency's (IEA) DSM Program contains tools to achieve the following:

- Reduce the demand peaks, especially when utilization of power comes close to its limits of availability.
- Fill the demand valleys where production is greater than consumption, to better utilize existing power resources.
- Reduce overall demand (strategic saving) in the context of delivering the required energy services by use of less energy (and not a reduction in services).
- Provide strategic growth especially to shift between one type of supply to another with more favorable characteristics, for example, in terms of the environment.¹²

In addition to energy efficiency measures such as DSM, as well as using non-CO₂ emitting energy sources (e.g., solar, geothermal, and wind) to reduce CO₂ emissions, fuel switching may also be employed. Fuel switching consists of replacing high-CO₂ emitting

fuel sources with biofuels such as plants and wood waste. Consumption of biofuels is considered to result in zero net emissions of CO₂. After DSM and renewable energy technologies, there are three main processes that may be employed to produce power from fossil fuel and then capture the CO₂ emissions: post-combustion control, pre-combustion control, and oxy-combustion. Figure 5 shows schematics of the processes to produce power and remove the resulting CO₂ emissions. Table 2 highlights different CO₂ separation techniques and their future potential.



Note: Figure accessed from International Energy Agency web site¹³

Figure 5 Three main processes that may be employed to produce power from coal and then capture the CO₂ emissions

Table 2 Current and future technologies for CO₂ capture

Capture method	Post-combustion decarbonisation CO ₂ /N ₂		Pre-combustion decarbonisation CO ₂ /H ₂		Oxyfuel conversion O ₂ /N ₂	
	Current	Future	Current	Future	Current	Future
Membranes	Polymeric	Ceramic facilitated transport Carbon molecular sieve	Polymeric	Ceramic Palladium Reactors Contactors	Polymeric	Ion-transport facilitated transport
Solvents/ Absorption	Chemical solvents	Improved process design Improved solvents Novel contacting equipment	Chemical solvents Physical solvents	Improved process design Improved solvents Novel contacting equipment	NA	Bio-mimetic solvents
Cryogenic	Liquefaction	Hybrid process Anti-sublimation	Liquefaction	Hybrid process	Distillation	Improved distillation
Solid Sorbents	Zeolites Activated carbon	Carbonates Carbon based sorbents	Zeolites Activated carbon Alumina	Dolomites Hydrotalcites Zirconates	Zeolites Activated carbon	Carbonates Hydrotalcites Silicates

Note: Table accessed from International Energy Agency web site¹³

For the post-combustion control process, capture of CO₂ in the combustion flue gas is typically performed with chemical absorption and subsequent solvent regeneration.¹³ Most existing CO₂ capture systems utilize this absorption-based technology.¹³ In chemical absorption, a large amount of energy is required to break the solvent-CO₂ bonds in the regeneration step: consumption of low-pressure steam for this step is equivalent to an estimated 0.178 kWh/kg (0.081 kWh/lb) CO₂ captured for 90% recovery (using a proprietary solvent similar to MEA developed and tested by Mitsubishi in 2002).¹³ Steam consumption for regeneration is projected to be as low as 0.083 kWh/kg (0.038 kWh/lb) CO₂ by 2015.¹³ Chen *et al* estimated that employing the standard MEA process at a standard air-blown pulverized coal (PC) plant decreases the net efficiency of power generation from 38% to 27.2%.¹⁴ Thus the primary challenge for the solvent-based systems is to recover the CO₂ with a minimum energy penalty. Improvements to the solvent type and process design are being investigated towards this end.¹³

In addition to chemical absorption, other principles of separation for post-combustion CO₂ capture include physical absorption with solvents, as well as the use of solid adsorbents, membranes, and cryogenics. Physical absorption with solvents is not suitable

for CO₂ capture from post-combustion flue gases which have a low CO₂ partial pressure (e.g., 0.1 atm, or 1.5 psia). In physical absorption there is a weaker solvent-CO₂ bond compared to chemical absorption, requiring a higher partial pressure of CO₂ for bonding. The CO₂ absorption capacity of a physical solvent increases almost linearly with increasing CO₂ concentration in the gas and can be higher than that of a chemical solvent at high CO₂ concentrations. Therefore, physical absorption is preferred over chemical absorption for the pre-combustion (post-gasification) process where the separation process operates at higher partial pressure of CO₂ (e.g., 10-15 atm, or 147-220 psia).

Use of solid adsorbents to capture CO₂ is a method similar to solvent absorption, and can operate at total pressures and temperatures higher or lower than liquid solvents. For post-combustion CO₂ capture, the solids are prone to attrition, be covered by surface deposits, or suffer chemical poisoning during the cyclic sorption-regeneration process. This leads to decreased reactivity and selectivity and thus less operating cycles – up to two orders of magnitude in some cases – compared to solvent absorption.¹⁵ As with solvent absorption, improvements in the solid sorbent type and process design are being investigated.¹⁶ The status of adsorption technology is discussed further below after the discussion on oxy-combustion.

A membrane separation process is based on the differences in physical or chemical interactions between gases and the membrane material. The factors of selectivity and permeability of the membrane allow one flue gas component to pass through the membrane faster than others. Multiple stages of the membrane process may be required to achieve a high purity CO₂ stream, which would increase energy consumption.¹⁴

The cryogenic process utilizes component volatility under cryogenic conditions, and separates CO₂ by condensing and cooling. Compression and cooling of the flue gas to a total pressure of 300 atm (4,409 psia) and -56 °C are required for 90% recovery of CO₂ at a conventional PC power plant, which also leads to increased energy consumption.¹⁴

The pre-combustion control process yields increased CO₂ concentration and fuel gas pressure compared to post-combustion control. These conditions allow for use of physical absorption with solvents or sorbents having weaker interaction with CO₂ relative to chemical solvents, and lower parasitic energy consumption from the CO₂ separation

process. The basic process includes fuel gasification and a water-gas-shift reaction to produce CO₂ and hydrogen. The CO₂ can then be removed by physical absorption or adsorption.

The oxy-combustion process uses concentrated oxygen instead of air for fuel combustion, which increases the concentration of CO₂ in the flue gas from 15% to 80% by volume (dry basis) or higher, thereby simplifying post-combustion purification.¹⁴ A power plant employing this type of technology requires an energy-intensive cryogenic process to produce the required concentrated oxygen, which leads to a net efficiency of power generation almost 30% less than that of a standard air-blown PC plant (i.e., 38% down to 27%).¹⁴ Advances in concentrating oxygen in air separation processes could improve the overall plant efficiency and make oxy-combustion a more attractive option.

Recent research has targeted solvent absorption and membrane separation for post-combustion CO₂ capture.¹³ The solid adsorption method has potential to be competitive with these technologies regarding energy conversion efficiency for the overall plant. A variety of solid sorbents are currently being examined for adsorption.¹⁶ These sorbents are classified into two types of adsorption processes: (1) cool-gas adsorption (40 – 60 °C), and (2) warm- or hot-gas adsorption (above 60 °C). Among the cool-gas adsorbents, simple porous solids such as activated carbons and zeolites are likely not suited well to post-combustion capture of CO₂, with respect to CO₂ adsorption capacities and CO₂/N₂ selectivities, along with the need for expensive pressure swing adsorption processes.¹⁶ Research results for functionalized porous sorbents, such as immobilized amine sorbents, indicate that porosity reduction by pore-filling of the functional groups offset the increase in CO₂ capacity by the functional groups.¹⁶ Among the warm- and hot-gas adsorbents, metal carbonate sorbents are an attractive option because they exist naturally and are low-cost (the cost of calcium carbonate, CaCO₃, in 2009 was \$21.0/short ton, or \$23.1/metric ton).¹⁷ One potential challenge is the need for fresh adsorbent to make up for decay in the active sorbent during the adsorption-desorption cycling. Retaining a capacity of 20-30% over 30 cycles would still be higher than achieved for simple and functionalized porous adsorbents.¹⁶

The three common types of metal carbonate sorbents are CaCO_3 , magnesium carbonate (MgCO_3), and sodium carbonate (Na_2CO_3). When comparing the CO_2 capture processes using each of these adsorbents, one important factor to consider is how well each process integrates with the plant steam cycle. A general figure showing the relationship of steam conditions to utilizable heat for electricity is provided in Appendix A. This figure shows that as the temperature and pressure of steam increase, the utilizable heat for electricity also increases. For example, at steam conditions greater than 500 psia (34.0 atm) and 524 °C, the utilizable heat for electricity is 31%, whereas the utilizable heat for electricity is reduced to 26% for steam at 300 psia (20.4 atm) and 400 °C (Appendix A). The adsorption/desorption temperatures for CaCO_3 , MgCO_3 , and Na_2CO_3 are 650 °C /900 °C, 275 °C /400 °C, and 60 °C /120 °C at a partial pressure of CO_2 (P_{CO_2}) = 0.01 atm (0.15 psia) for adsorption and P_{CO_2} = 1 atm (14.7 psia) for desorption, respectively (Appendix A). The equilibrium curve for CaO is also provided in Appendix A. When the metal carbonate capture process is integrated with the overall plant steam cycle, there is a higher percentage of utilizable heat for electricity provided by the steam generated from the recovery of heat of adsorption at higher temperatures. For the CO_2 capture process using CaCO_3 , high-pressure (2,415 psia, or 164 atm), high-temperature (540 °C) steam can be generated from the exothermic reaction inside the adsorption reactor (650 °C). This additional steam can help offset the energy used for sorbent regeneration (see Section 2.2.2.2 below for details). However, MgCO_3 and Na_2CO_3 are not useful for heat recovery due to the low adsorption temperatures (275 °C and 60 °C, respectively). CaCO_3 is also a common, low-cost adsorbent (on the order of hundreds of dollars less per short ton, or hundreds of dollars less per metric ton, than MgCO_3 and Na_2CO_3). Hence, on the basis of minimizing energy loss and the cost of adsorbent, CaCO_3 was chosen for further investigation here to remove 90% of the CO_2 emissions at a coal-fired power plant.

There have been several studies that include energy use analyses related to the limestone adsorption-desorption process. The most closely related study – with respect to plant size and integration of the CO_2 capture process with the rest of the plant – was performed by Ramkumar *et al.*¹⁸ The study simulated energy use of a hydrated lime ($\text{Ca}(\text{OH})_2$) process with multiple cycles for 90% capture of CO_2 . A study by Grasa *et al* included an analysis of heat recovery from the hot CO_2 stream out of the desorption reactor.¹⁹ However, the

simulated process removed only 33% of the CO₂ in the flue gas stream. A study by the Department of Energy's National Energy Technology Laboratory (DOE/NETL) examined reactor design for CO₂ capture using different sorbents, and indicated that DOE is targeting a 30-50% reduction of the regeneration heat duty of the MEA process.²⁰ Sjoström has evaluated solid sorbents for CO₂ capture, and showed that metal carbonates consume a higher theoretical amount of regeneration energy than MEA (4,530 kJ/kg, or 1,947 Btu/lb CO₂ captured).²¹ The European Commission has also investigated the limestone adsorption-desorption process, and recovered the desorption heat in the steam cycle. However, only 60% of the CO₂ was captured from the boiler's flue gas.²² These select studies of limestone use in the adsorption-desorption process for CO₂ capture at an air-blown PC power plant indicate the need for energy use details – for the adsorption-desorption process itself, and the heat recovery benefit to the plant. A more detailed description of these studies is provided in Section 3.3.

1.5 Coal Costs

Apart from future environmental regulations, it is also worthwhile to evaluate the impacts of coal transportation cost and coal free-on-board (FOB) mine price on economic performance at power plants. FOB mine price indicates a sales transaction wherein the coal supplier makes the coal available for pick up at a specified port or terminal at a specified price, and the buyer then pays for the subsequent transportation and insurance. Combined transportation and FOB price was considered in the range of \$20.0 to \$32.0/short ton (\$22.0 to \$35.3/metric ton) of PRB coal delivered in 2006 – with the most probable cost at \$27.5/short ton (\$30.3/metric ton).²³ The equivalent delivered cost of Illinois coal – which uses Illinois #6 bituminous coal in this study – was considered in the range of \$30.0 and \$40.0/short ton (\$33.1 and \$44.1/metric ton), with \$35.0/short ton (\$38.6/metric ton) as the most probable cost.²³ The influence of ash disposal – which may be considered as part of the coal cost – on economic performance was not considered directly in this study; however, this cost is part of the PM control cost.

1.6 Introduction to Integrated Environmental Control Model

The Integrated Environmental Control Model (IECM) was developed by Carnegie-Mellon University for the DOE/NETL. The model was developed to provide preliminary

technical performance and cost estimates for new base load power plants as well as existing plants considering technology retrofits. IECM allows for specifying fuel cost, fuel composition, and environmental compliance limits for SO₂, NO_x, PM, mercury, and CO₂. The model also considers the type of air quality control technology (e.g., wet or dry flue gas desulphurization; electrostatic precipitator or fabric filter; and low-NO_x combustion modification and/or selective catalytic reduction). The model allows the user to configure a system of processes, set parameters for the overall system and each process, and generate various results of the system performance. The model has a module that allows inputting the coal characteristics (proximate and elemental compositions and heat value) and the delivered coal cost.

IECM is unique in combining boiler parameters and coal characteristics to estimate the capital cost of the environmental control systems. Aside from these unique features, this model has a refined retrofit factor analysis. In IECM, there are several cost sub-modules that can have different retrofit factors. The IECM cost sub-modules for FGD installation are: reagent feed, SO₂ removal, flue gas handling, solids handling, general support, and miscellaneous equipment. The sub-modules for selective catalytic reduction (SCR) installation are: reactor housing, ammonia injection, ducts, air pre-heater modifications, induced draft (ID) fan differential, structural support, and miscellaneous equipment. The sub-modules for PM control installation are: particulate collector, duct work, fly ash handling, and ID fan differential. The sub-modules for mercury control installation are: sorbent injection, sorbent recycle, duct work, sorbent disposal, and pulse jet fabric filter (FF). The sub-modules for CO₂ control installation are: SO₂ polisher/direct contact cooler, flue gas blower, CO₂ absorber vessel, heat exchangers, circulation pumps, sorbent regenerator, reboiler, steam extractor, sorbent reclaimer, sorbent processing, and drying and compression unit. Each of the sub-modules can have a different retrofit factor. For example, if a retrofit site has adequate land for a FGD installation but the FGD must be located at a remote location from the existing equipment, the IECM model can be adjusted to increase the retrofit factor for flue gas handling, and set a low to no retrofit factor for the remaining FGD areas. This option makes IECM very powerful in estimating the cost to retrofit a utility boiler.

In addition to capital cost flexibility, IECM allows the user to specify the site-specific reagent cost, by-product disposal cost or sale price, labor charge, and other costs. If these operating costs are not specified, then default values are assumed within IECM. There are also a number of IECM configurations available to evaluate the different methods of reducing emissions while complying with regulations. The control technologies available are listed in Table 3. The model uses all of these inputs to estimate the reagent usage, power consumption, by-product production, and electricity costs.

Table 3 Control technologies used in IECM

Parameter	Technology
<i>Combustion</i>	
Fuel type	Coal
NO _x	Low NO _x Burner (LNB), LNB + Over Fire Air, Gas Reburn, Selective Non-Catalytic Reduction (SNCR), or LNB + SNCR
<i>Post-Combustion</i>	
SO ₂	Wet Flue Gas Desulfurization, or Spray Dryer Absorber
NO _x	Hot-Side Selective Catalytic Reductions
PM	Cold-Side Electrostatic Precipitator, or Fabric Filter
Mercury	Activated Carbon Injection (ACI), or ACI + Water Injection
CO ₂	Amine System, or O ₂ -CO ₂ Recycle

The history of IECM dates back to the early 1980's. The model has been supported by the DOE/NETL under projects for PC combustion plants and integrated gasification combined cycle (IGCC) systems. The development of IECM began as a tool to assess the impact of fossil-fuel power plant emissions and pollution control options on acid rain. Post-combustion control options were added for SO₂, NO_x, and PM emissions from coal combustion systems, and the fully integrated tool contained a cost model linked to a performance model for each technology in the system. Studies performed in the 1980's using IECM focused on basic pollution control economics from fossil-fuel combustion systems, considering variables in the areas of technology availability, economics, and regulations – particularly the Acid Rain Program.^{24,25,26}

In the 1990's, combustion abatement technologies, as well as post-combustion control technologies, were added to IECM. Studies performed in this decade were influenced heavily by Phase I of the Acid Rain Program, which targeted mostly coal-fired electric utilities for SO₂ and NO_x reductions.^{27,28,29} In the 2000's, natural gas combined cycle and carbon capture and sequestration (CCS) were added to IECM, as well as a gasifier option.

Studies in this decade focused primarily on the performance of CO₂ capture technology.^{30,31} The Illinois State Geological Survey (ISGS) also performed an economic and environmental performance evaluation in 2007 of power plants in the Illinois Basin using IECM.³² The evaluation included CO₂ capture cost with electricity supplement and various control levels.

The main model outputs of interest in this study are \$/MWh values and capital and operation and maintenance (O&M) costs (fixed and variable) for environmental control units, boiler unit, and fuel, as well as for the entire plant. Additional details of the IECM modeling program can be found at the IECM website.³³

1.7 Objectives

The objectives of this research were the following:

1. Evaluate the cost to generate one megawatt-hour of electricity from firing bituminous coal, blends of bituminous and PRB coals, and PRB coal while complying with the Illinois SO₂, NO_x, PM and mercury emission limits for electric generators using IECM.
2. Assess factors that may influence the electric generator's fuel selection process, such as the fuel type – which includes the delivered cost, heat content, and chemical property of the fuel – and plant size.
3. Determine the impacts – economics, and emissions of SO₂, NO_x, PM and mercury – caused by adding CO₂ capture with a MEA process on results from the IECM evaluation with no CO₂ capture.
4. Evaluate the limestone carbonation-calcination (LSCC) process using ChemCAD, based on energy use, emissions of SO₂, NO_x, PM and mercury, and comparison of the LSCC process with the MEA process to remove CO₂ from flue gas streams.

1.8 Research Justification

The IECM portion of this study may be used by managers of coal-fired power plants to assist in determining the most cost-effective approach when burning select coals while simultaneously meeting stack emission regulations. Law- and policy-makers may use results from this study to evaluate mining options to lower FOB mine costs, evaluate coal

transportation costs, and/or develop policy options concurrent with desired impacts on coal production and sales. Results from the ChemCAD portion of this study can be useful for assessing select CO₂ control technologies regarding impact on energy conversion efficiency.

2 METHODS

2.1 IECM Evaluation of COE

IECM version 5.2.1 (2007) was used to evaluate the impacts of environmental compliance costs and delivered coal cost on the cost of electricity (COE) at a coal-fired power plant.³³ The COE reflects the plant cost levelized over the plant lifetime and size, and is presented as \$/MWh. Environmental compliance costs consist of capital (e.g., equipment) and fixed and variable O&M costs, such as labor and materials. Delivered coal cost contributes to the variable O&M costs of the power plant.

2.1.1 Coal Specifications

The typical coal specifications for Illinois #6 bituminous (hereafter called “Illinois” or “IL”) and PRB were based on the fuel database built in IECM, and are listed in Table 4. The coal specifications of the 70/30 PRB/IL blend were the weighted-averages of the two individual coals according to their ratio in the blend. In addition to existing power generators, coal specifications for new mine-mouth coal-fired boilers were also developed based on interviews with power generators.²³

Table 4 Coal specifications for the scenarios tested

Property	Coal type ^a			
	Illinois	Western PRB	70/30 PRB/IL ^b	Illinois mine-mouth
HHV, Btu/lb ^c (kJ/kg)	10,900 (25,353)	8,340 (19,399)	9,108 (21,185)	9,500 (22,097)
Carbon, wt %	61.353	48.214	52.156	51.998
Hydrogen, wt %	4.211	3.312	3.582	3.568
Oxygen, wt %	6.035	11.878	10.125	5.115
Chlorine, wt %	0.170	0.010	0.058	0.144
Sulfur, wt %	3.000	0.300	1.110	4.000
Nitrogen, wt %	1.163	0.700	0.839	0.986
Ash, wt %	11.028	5.324	7.035	23.144
Moisture, wt %	13.033	30.261	25.093	11.045
Mercury, ppm _m ^d	0.090	0.100	0.097	0.090
^a Significant figures for reported values are consistent with those reported by the IECM model ^b 70/30 PRB/IL = Mix of 70% PRB coal and 30% Illinois coal by weight ^c HHV = Higher Heating Value ^d ppm _m = Parts per million by mass				

2.1.2 Environmental Control Configurations

Coal-fired power plants are expected to comply with emission regulations by switching to a lower sulfur fuel, installing air quality control technologies, or both fuel switching and environmental control technology installations. There are a wide range of IECM configurations available to evaluate the different methods of reducing emissions to comply with emission regulations. Figure 6 shows the environmental control configurations for the coals used in this study. The MEA process is included only for scenarios with 90% CO₂ capture. Details for each environmental control configuration and the rationale for environmental compliance are described in the following subsection.

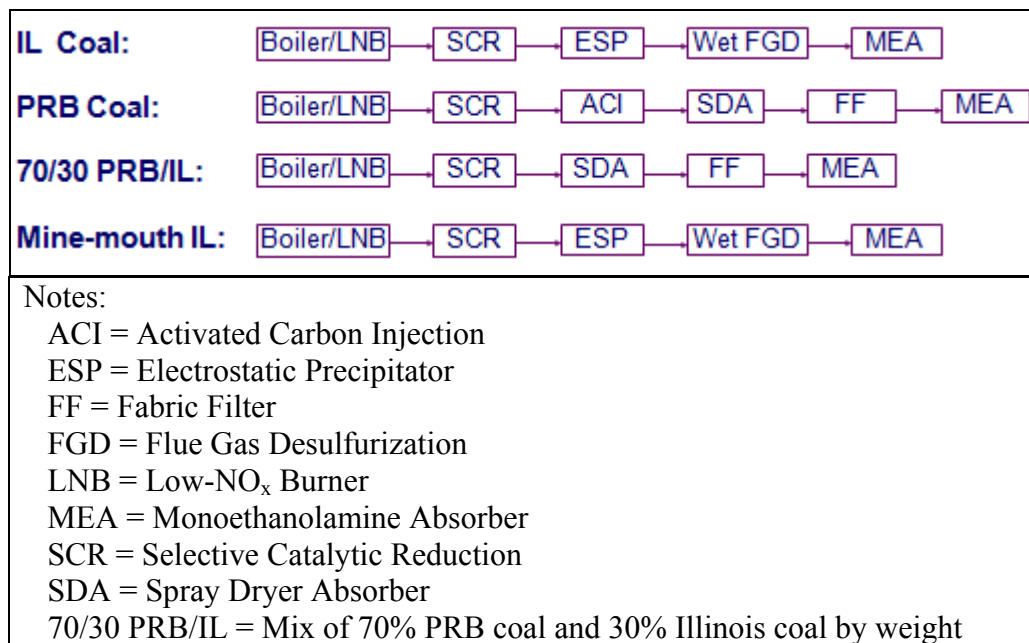


Figure 6 Environmental control configurations for different coals

2.1.3 Scenarios

Assumptions for particular IECM performance requirements were developed and used as inputs in IECM. These inputs were used to evaluate scenarios with different coal feed types and environmental controls. Table 5 shows model parameters and values for select coal scenarios.

Table 5 Model parameters and values for scenarios

Model parameter	Illinois		PRB		70/30 PRB/IL ^a		Illinois mine-mouth (new plant)	
	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture
Fuel cost (total as-delivered), \$/short ton (\$2007)	Low: 30.0 Mid: 35.0 High: 40.0	Low: 30.0 Mid: 35.0 High: 40.0	Low: 20.0 Mid: 27.5 High: 32.0	Low: 20.0 Mid: 27.5 High: 32.0	All combinations of PRB coal costs with IL coal costs	All combinations of PRB coal costs with IL coal costs	Low: 17.6 High: 25.0	Low: 17.6 High: 25.0
Fuel cost (total as-delivered), \$/metric ton (\$2007)	Low: 33.1 Mid: 38.6 High: 44.1	Low: 33.1 Mid: 38.6 High: 44.1	Low: 22.0 Mid: 30.3 High: 35.3	Low: 22.0 Mid: 30.3 High: 35.3	All combinations of PRB coal costs with IL coal costs	All combinations of PRB coal costs with IL coal costs	Low: 19.4 High: 27.6	Low: 19.4 High: 27.6
Boiler size, MW _e	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650	Small: 175 Large: 650
SO ₂ emission limit, lb/mmBtu	0.25	< 0.25	0.25	< 0.25	0.25	< 0.25	0.25	< 0.25
SO ₂ emission limit, kg/MJ	1.1x10 ⁻⁴	< 1.1x10 ⁻⁴	1.1x10 ⁻⁴	< 1.1x10 ⁻⁴	1.1x10 ⁻⁴	< 1.1x10 ⁻⁴	1.1x10 ⁻⁴	< 1.1x10 ⁻⁴
SO ₂ control technology	Wet FGD	Wet FGD	SDA	SDA	SDA	SDA	Wet FGD	Wet FGD
NO _x emission limit, lb/mmBtu or % removal ^b	0.11 or 80%	< 0.11 or < 80%	0.11 or 80%	< 0.11 or < 80%	0.11 or 80%	< 0.11 or < 80%	0.11	< 0.11
NO _x emission limit, kg/MJ or % removal ^b	4.7x10 ⁻⁵ or 80%	< 4.7x10 ⁻⁵ or < 80%	4.7x10 ⁻⁵ or 80%	< 4.7x10 ⁻⁵ or < 80%	4.7x10 ⁻⁵ or 80%	< 4.7x10 ⁻⁵ or < 80%	4.7x10 ⁻⁵ or 80%	4.7x10 ⁻⁵ or 80%
NO _x control technology	LNB + SCR	LNB + SCR	LNB + SCR	LNB + SCR	LNB + SCR	LNB + SCR	LNB + SCR	LNB + SCR
PM emission limit, lb/mmBtu	0.03	0.03	0.03	0.03	0.03	0.03	0.015	0.015
PM emission limit, kg/MJ	1.3x10 ⁻⁵	1.3x10 ⁻⁵	1.3x10 ⁻⁵	1.3x10 ⁻⁵	1.3x10 ⁻⁵	1.3x10 ⁻⁵	6.4x10 ⁻⁶	6.4x10 ⁻⁶
PM control technology	ESP	ESP	FF	FF	FF	FF	ESP	ESP

Table 5 (cont.)

Model parameter	Illinois		PRB		70/30 PRB/IL ^a		Illinois mine-mouth (new plant)	
	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture	No CO ₂ capture	CO ₂ capture
Mercury emission limit, lb/MWh or % removal	90%	90%	ACI	ACI	90%	90%	2.1x10 ⁻⁵	2.1x10 ⁻⁵
Mercury emission limit, kg/MWh or % removal	90%	90%	ACI	ACI	90%	90%	9.5x10 ⁻⁶	9.5x10 ⁻⁶
Mercury control using ACI, lb/10 ⁶ acf ^{c,d}	N/A	N/A	5	5	N/A	N/A	N/A	N/A
Mercury control using ACI, kg/10 ⁶ acm ^{c,d}	N/A	N/A	8.0	8.0	N/A	N/A	N/A	N/A
CO ₂ emission limit, % removal ^d	N/A	90%	N/A	90%	N/A	90%	N/A	90%
CO ₂ control technology ^d	N/A	MEA	N/A	MEA	N/A	MEA	N/A	MEA
^a 70/30 PRB/IL = Mix of 70% PRB coal and 30% Illinois coal by weight ^b 80% was the operational value, as it resulted in lower NO _x emissions ^c Activated Carbon Injection (ACI) at 5 lb/10 ⁶ acf (8.0 kg/10 ⁶ actual cubic meter, or acm) is an alternative compliance option to the % removal requirement (see Table 1) ^d N/A = Not Applicable								

For the different scenarios, the Illinois mine-mouth plant as a new source followed the USEPA emission standards for SO₂, NO_x, PM and mercury, and other plants followed the 2015 Illinois standards (see Table 1). Parameters tested were: coal type and cost; control technology for SO₂, NO_x, PM and mercury; boiler type; and boiler size. Tests were also performed with and without 90% CO₂ capture from the flue gas. The breakeven costs of electricity with respect to use of different fuels, as well as the capital and O&M costs, were estimated for the different scenarios. Unless noted, default values that are provided with IECM were used within IECM for process and cost assumptions.

2.1.3.1. Scenario 1 – SO₂, NO_x, PM and Mercury Controls with no CO₂ Capture

The delivered coal costs for PRB and Illinois coals cover a range of costs to Illinois plants: \$30.0, \$35.0, and \$40.0/short ton (\$33.1, \$38.6, and \$44.1/metric ton) for Illinois coal; \$20.0, \$27.5, and \$32.0/short ton (\$22.0, \$30.3, and \$35.3/metric ton) for PRB coal; and \$17.6 and \$25.0/short ton (\$19.4 and \$27.6/metric ton) for Illinois mine-mouth coal. Additional cost for blending of coals was not considered. Possible boiler types include tangential- and wall-fired pulverized coal (PC), as well as cyclone boilers. Tangential is most common and was therefore used for this study. The retrofit factors for different processes of the plant were determined by first testing a 1,200 MW_e plant with PRB coal, and adjusting the retrofit value until the combined capital cost for the SDA and FF units was \$400 MM. This cost corresponds to the value reported by the utility companies that were interviewed as part of this project.²³ The resulting retrofit value was 3.0, and this factor was then applied to the wet FGD, SDA, SCR, and FF units for all cases. This retrofit factor represents the impacts of the inflation and market prices change since the year the cost models used in IECM (version 5.2.1, 2007) were developed. Therefore, the retrofit factor was used as a correction of cost change between those in IECM and the 2007 market costs, rather than only used to represent the difficulty level of retrofitting. A retrofit value of 1.0 was used in cases where the ESP continued to operate without any retrofitting.

For SO₂ control, wet FGD was used when burning Illinois coal, due to its high-sulfur content. Similarly, wet FGD was used in the Illinois mine-mouth case. ACI was used only for 100% PRB coal because the coal's chlorine content (0.01 wt %) is significantly

lower than Illinois coal (0.17 wt %). The lower amount of chlorine leads to less oxidized mercury and less mercury capture in the SDA/FF system. When burning 100% Illinois coal, the higher chlorine content, along with the inherent SCR oxidation of elemental mercury to oxidized mercury and the inherent oxidized mercury capture by the wet FGD system, is sufficient to achieve 85-90% mercury removal.³⁴ It has also been demonstrated for a selective catalytic reduction - spray dryer absorption/fabric filtration (SCR-SDA/FF) configuration that a 93/7 (wt %) PRB/IL coal blend removes 50% mercury, and a 86/14 (wt %) PRB/IL coal blend removes 80% mercury.³⁵ The coal blend ratio used in this study was 70/30 (wt %) PRB/IL and is expected to remove a minimum of 90% mercury with a SCR-SDA/FF configuration. For the Illinois mine-mouth scenario, a higher mercury removal of 98% is necessary to comply with the 2.1×10^{-5} lb/MWh gross (9.5×10^{-6} kg/MWh gross) requirement in Table 1.

2.1.3.2. Scenario 2 – SO₂, NO_x, PM and Mercury Controls and 90% CO₂ Capture with MEA

An MEA process was added to the IECM scenarios without CO₂ capture, to capture 90% CO₂ from the flue gas. For the MEA process, a maximum SO₂ concentration of 10 parts per million by volume (ppm_v) into the amine scrubber is recommended in order to avoid excessive contamination of the solvent.³⁶ The wet FGD module built in IECM is capable of achieving a minimum SO₂ concentration to less than 50 ppm_v for the 100% Illinois, 100% PRB, and 70/30 PRB/IL blend fuel options, but only 132 ppm_v for the Illinois mine-mouth scenario. Therefore, 132 ppm_v was used as the standard limit after the wet FGD across all four fuel options, in order to obtain the most consistent comparison possible within IECM. A retrofit factor of 1.0 was used for the MEA process units. Issues pertaining to transportation and sequestration of CO₂ were not included in this study.

2.2. Assessment of Limestone Carbonation-Calcination Process

2.2.1 Description of Reference Plants

The reference (hereafter referred to as “base”) power plant for this study is a conventional, air-blown PC power plant based on a U.S. DOE-funded study.³⁷ A schematic of the base plant is shown in Figure 7.

The base plant with an MEA unit added for CO₂ capture (Figure 8) uses the same typical process components and post-combustion control units as the base plant. An assessment of the energy and cost performances of this case was performed by Chen *et al* and is described below.¹⁴

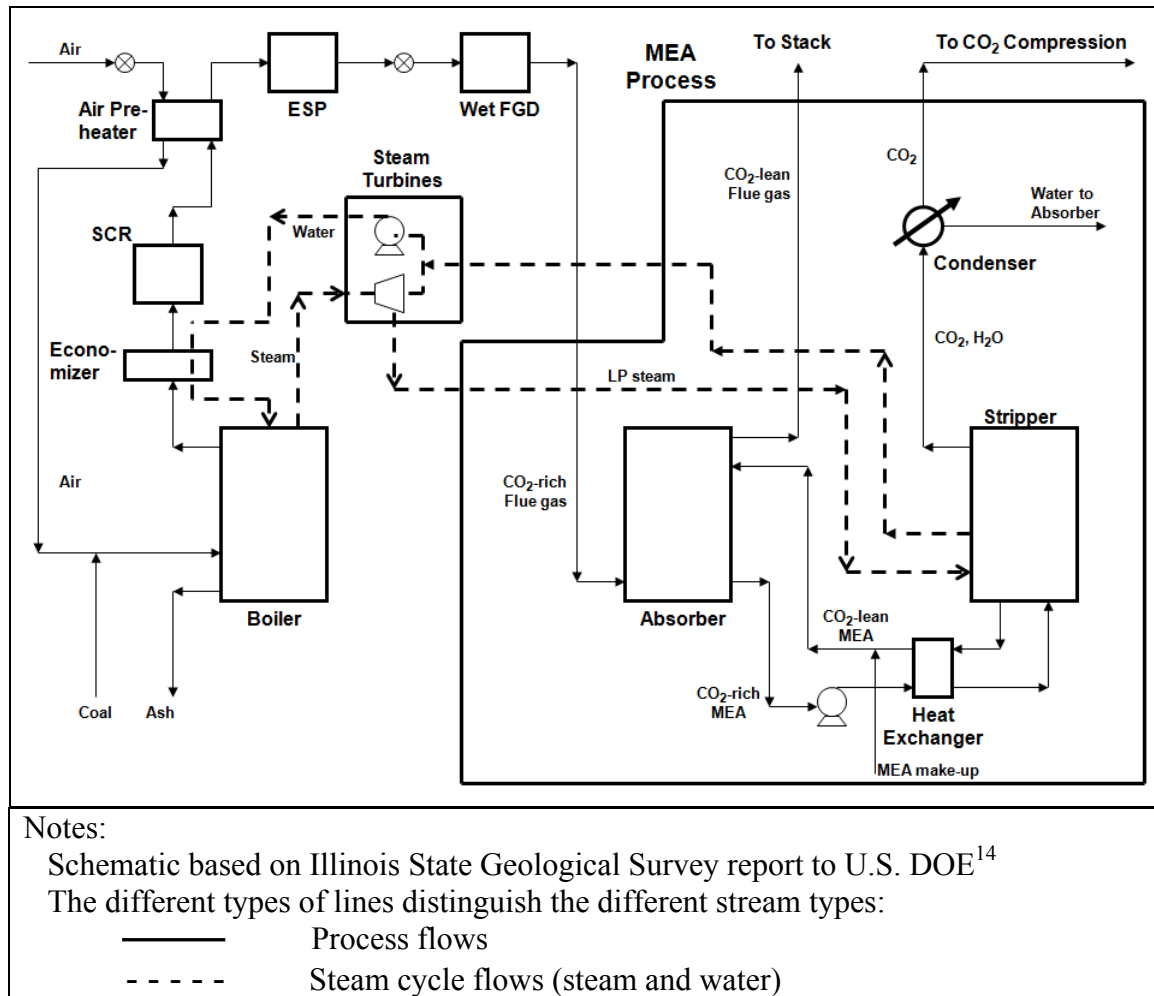


Figure 8 Schematic of reference power plant with MEA process.

The CO₂-rich flue gas from the FGD unit flows upward through the absorber unit, where the MEA solution absorbs CO₂ from the boiler flue gas. The total pressures into and out of the absorber were 17.4 and 14.4 psia (1.18 and 0.98 atm) respectively, while the temperature of the flue gas was about 50 °C.¹⁴ The liquid flow rate in the absorber was based on a 5:1 liquid-to-gas (L/G) molar ratio (MEA concentration of 30 wt % in water) to achieve 90% CO₂ removal.¹⁴ The L/G ratio is important because a greater amount of MEA liquid requires the consumption of more energy by the stripper.

The CO₂-laden MEA solution from the absorber is transported to the stripper for regeneration. Super heated low-pressure steam – at 60 psia (4.1 atm) and 246 °C – is used to meet the requirements of regeneration temperature.¹⁴ The CO₂ gas stream out of the stripper was 25 psia (1.7 atm) and 95 °C, with the inlet and outlet liquid streams at 95 °C and 116 °C in order to drive the release of CO₂.¹⁴ The pressures of the liquid streams into and out of the stripper were both 25 psia (1.7 atm).¹⁴ Loss of the pure MEA solution occurs due to stoichiometric reactions with acid gases in the incoming flue gas to the absorber, as well as nominal loss (see Appendix B2). As shown in Figure 8, a condenser is required to cool down the CO₂ stream from the stripper and remove water before the CO₂ compressor.

The operating conditions of the plant steam cycle were modified slightly from the base plant in order to accommodate the super heated low-pressure steam – at 246 °C and 60 psia (4.1 atm) – required at the stripper. Additional summary information for the plant steam cycle is provided in Appendix B2. A summary table of the operating equipment and parameters of the base plant with MEA is shown below in Section 2.2.2.2.

2.2.2. Limestone Carbonation-Calcination Process

The process of using a metal-oxide base for CO₂ capture is termed “carbonation-calcination.” The general reaction for a metal “Me” with CO₂ is shown in Equation 1.



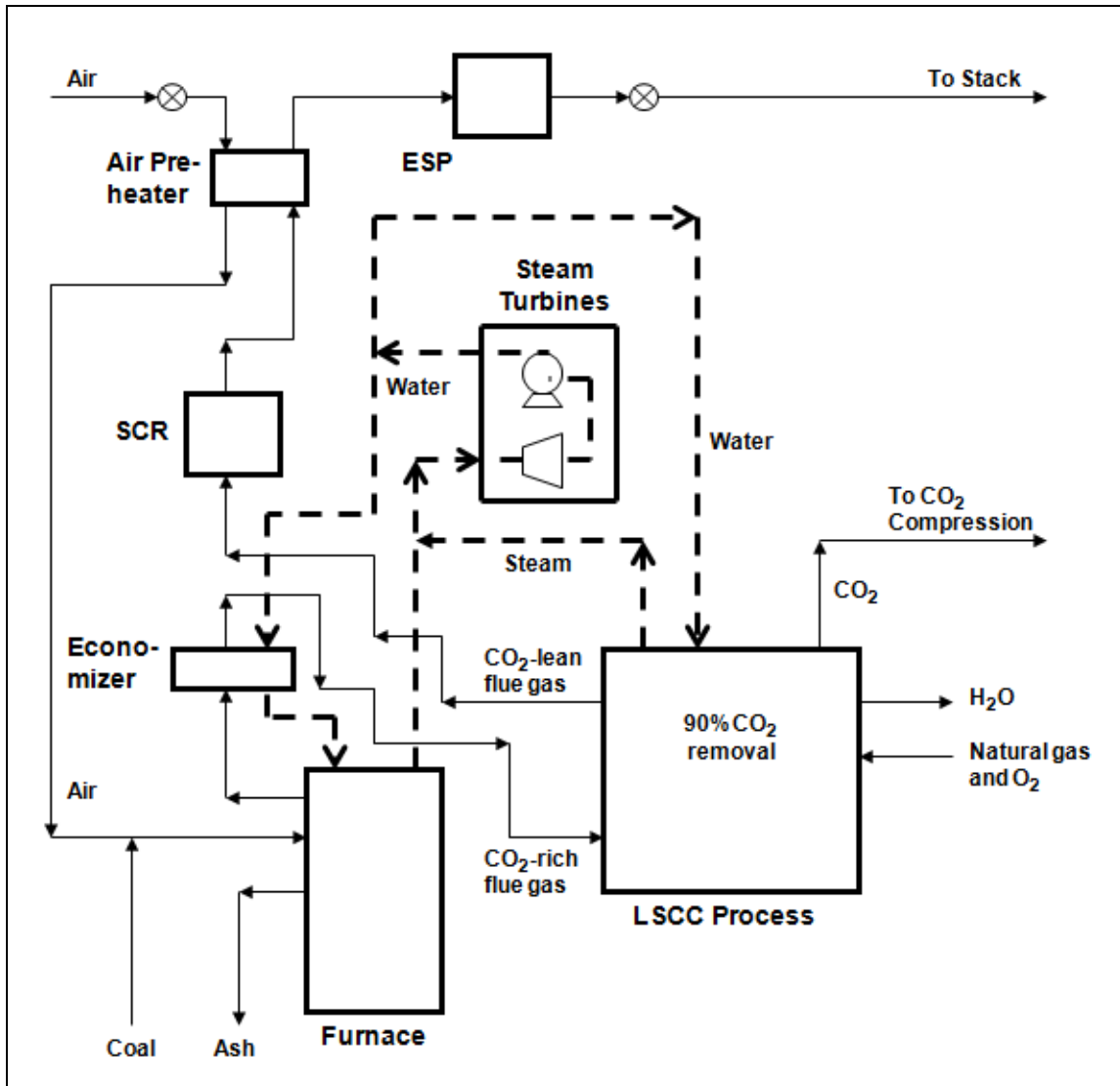
Knowledge of the required reactions for this process of CO₂ separation dates back to 1867.³⁸ In 1967, a fluidized bed pilot plant using 40 short tons/day (36 metric tons/day) CaO was demonstrated to capture CO₂ for the production of pipeline gas.³⁹ The process is similar to the MEA process shown above in Figure 8, with the major difference being that a solid sorbent is utilized to capture CO₂ instead of a liquid solvent. As with MEA, the solid sorbent is assumed to possess a certain degree of regenerability. The CO₂-rich flue gas first reacts with the solid sorbent in the “carbonator” (similar to the absorber in Figure 8). The reacted sorbent is then transported to the “calciner” (similar to the

regenerator in Figure 8). In the calciner, the sorbent is regenerated and the captured CO₂ is released for further processing. The use of a solid instead of a liquid solvent introduces the need for an alternative transportation method between the two units. This transportation method is discussed below in Section 2.2.2.2.

2.2.2.1 Overall Description of Power Plant with LSCC Process

The design and throughput for the plant boiler used in this study was based on the base (DOE reference) plant.³⁷ The properties of the bituminous coal used to fire the boiler are provided in Table C.1 in Appendix C. Modifications were made to the design and configuration of the base plant in order to integrate the limestone carbonation-calcination (LSCC) process for capturing CO₂ within the boiler flue gas. These modifications are described in more detail below in Section 2.2.2.2. Several parameters used for the power plant in the DOE study are also used for the plant in this study: plant size of 533 MW_e gross output; use of bituminous coal as boiler fuel; and use of SCR and ESP units.

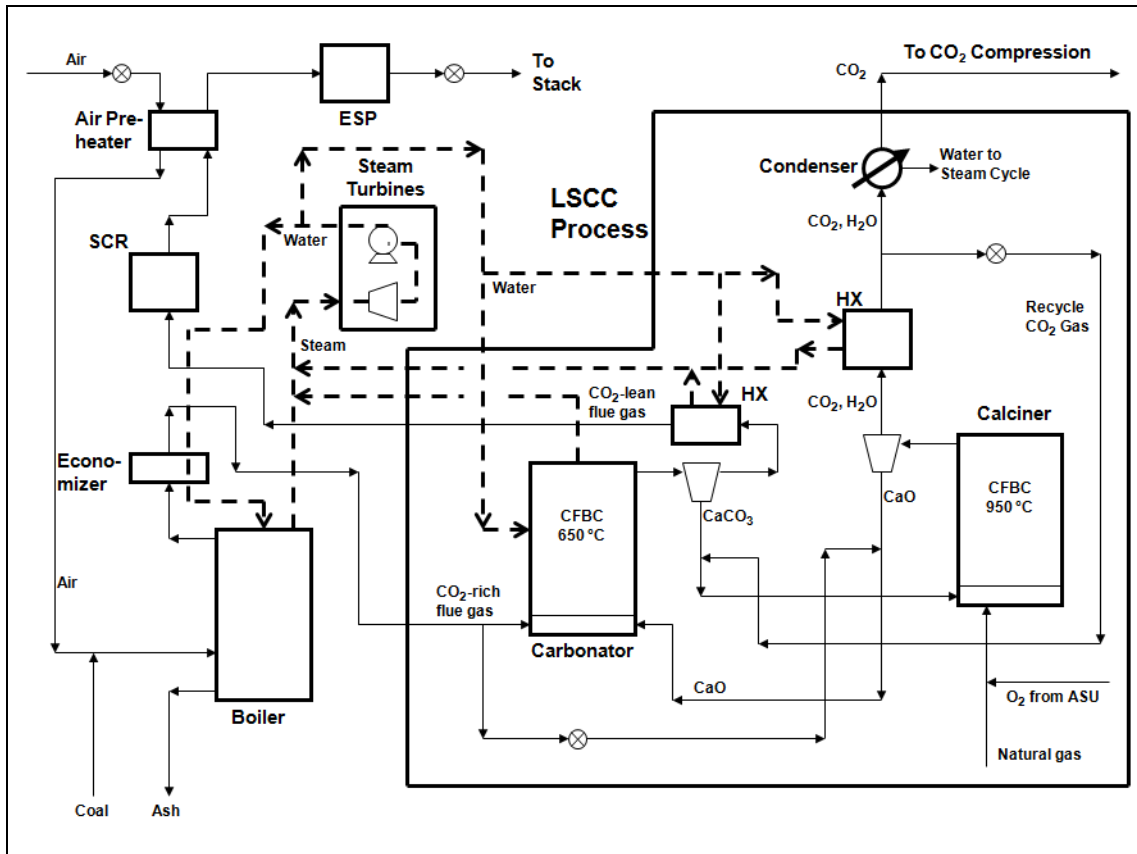
Figures 9 and 10 show the power plant configuration with a CO₂ capture system utilizing the LSCC process developed in this study.



Note: The different types of lines distinguish the different stream types:

- Process flows
- Steam cycle flows (steam and water)

Figure 9 Schematic of overall power plant with LSCC process



Notes:

ASU = Air Separation Unit

CFBC = Circulating Fluidized Bed Carbonator/Calciner

HX = Heat Exchanger

The different types of lines distinguish the different stream types:

———— Process flows

----- Steam cycle flows (steam and water)

Figure 10 Schematic of LSCC process

Developing and assessing the LSCC process consisted of the following steps:

- 1) Select the chemical compounds for inputs and desired outputs for the overall system (utilizing given conditions from the base power plant);
- 2) Estimate the appropriate process conditions for the LSCC system integrated with the base power plant;
- 3) Set up the appropriate unit operations to yield the desired outputs; and
- 4) Simulate the system.

Results from step 4 yield values for energy use or supply by different unit components of the system, which can then be compared to a similar-size power plant using an MEA process to capture CO₂. This is an ideal, best-case study for LSCC, intended as a proof-of-concept study. A primary goal of this study is to make a simplified process design that is sufficient to include major unit operations for assessing the overall energy use performance for a retrofitted power plant. The process is assessed under an ideal condition: stoichiometric adsorption reaction between CO₂ and CaO; no sorbent attrition loss or degradation; and maximum heat transfer efficiency. Additional explanations and data sources for the assumptions made below for the LSCC process are provided in Appendix C.

2.2.2.2 Detailed Process Design for LSCC Process

Carbonator:

A post-boiler, pre-SCR flue gas at 340 °C is introduced into a mixing chamber (indicated by a box at the bottom of the “Carbonator” in Figure 10 immediately before the carbonator. This flue gas mixes with the hot CaO (950 °C) in order to obtain a flue gas/CaO mixture inlet temperature to the carbonator near 650 °C. Additional details about the mixing process are provided below in this section. The CO₂-rich flue gas is withdrawn downstream of the boiler, after the economizer and before the SCR, at 14.5 psia (0.99 atm). This section of the flue duct is retrofitted with additional ducting to and from the carbonator. The CO₂-lean flue gas exiting from the carbonator is then cooled down back to 340 °C using an additional water-gas heat exchanger, and returns to the pre-SCR duct at 12.2 psia (0.83 atm). The heat exchanger is integrated with the plant steam cycle and produces high-pressure (HP), high-temperature steam at 540 °C and 2,415 psia (164 atm), based on a simulated amount of feed water (Section 2.2.3). This configuration for heat recovery is intended for energy calculation purposes, and the practical configuration of the heat exchanger may be different. Similar studies do not have details about this specific part of the process configuration.^{40,41,42}

A fluidized bed reactor (FBR) was selected as the type of reactor for the carbonator, because it is capable of handling a large amount of solid material in a continuous operation. The FBR also favors gas solid reactions by providing efficient gas-solid

mixing, and fresh surface exposure due to attrition of the solid. FBRs have been widely used for over 25 years, and were used for similar studies with LSCC processes.^{14,39} Entrained bed and fixed bed reactors were also considered for this study. These reactor types were dismissed due to limitations in flow capacity, gas-solid mixing, and heat transfer efficiency. Reactor sizing is beyond the scope of this study.

The temperature in the carbonator was maintained at 650 °C in order to achieve the desired 90% CO₂ capture in the flue gas. This temperature is assumed to be reasonable based on the CaCO₃/CO₂ equilibrium curve (Appendix A1) for a CO₂ partial pressure of 0.2 psia (0.014 atm), which is equivalent to 90% CO₂ capture from the inlet flue gas with CO₂ partial pressure of 2 psia (0.14 atm). Since the adsorption of CO₂ by lime to produce limestone is an exothermic reaction, the temperature was held constant by using a heat exchanger (boiler tube bundle) within the unit for water-gas heat exchange. This heat exchanger generates HP, high-temperature steam at 540 °C and 2,415 psia (164 atm), based on a simulated amount of feed water (Section 2.2.3). A limestone particle diameter of 0.5 mm was used for this study, consistent with a similar study by Shimizu.⁴² The gas-solid heat transfer efficiency between the lime particles and boiler flue gas was assumed to be 100%. A calculation for the pressure drop required in the carbonator (Appendix C3), which provides fluidization of the particles, shows that 0.10 atm (1.47 psia) is reasonable. This pressure drop assumes a reactor height of 30 m.

As shown in Figure 10, CaO at 950 °C from the calciner is combined with the post-boiler flue gas in a mixing chamber before the carbonator. The required temperature of the boiler flue gas was calculated to be 353 °C, based on lowering the hot CaO from 950 °C to a gas-solid mixture temperature of 650 °C. The calculation considers the heat capacities and mass flow rates of the flue gas and CaO streams, and is provided in Appendix C3. The flue gas temperature after the economizer and before the SCR is 340 °C – for practical purposes, this temperature is employed in the process simulation to provide the required flue gas/CaO mixture temperature within 10 °C of 650 °C.

The CaO flow rate in the carbonator is based on a 1:1 mole ratio of Ca to CO₂. This is an ideal, best-case condition, assumed for this proof-of-concept study. This ratio is important because a larger amount of limestone solids is associated with more cooling

required at the carbonator and additional heat duty required at the calciner, as well as greater difficulty circulating the solids in the adsorption and desorption loop. The study by Ramkumar *et al* used a calcium hydroxide ($\text{Ca}(\text{OH})_2$) sorbent at 1.3 (mol) CaO/CO_2 , for a 90% CO_2 capture process.¹⁸ Abanades used 3:1 (mol) CaO/CO_2 , along with a fresh makeup feed of 1:10 (mol) $\text{CaCO}_3/\text{CO}_2$, for a 90% CO_2 capture process similar to this study.⁴¹ Another estimate, by Statnik, reported a complete loss of active CaO after 12 cycles.⁴³ The decay in carbonation conversion in Statnik's estimate was attributed to textural changes due to sintering, which include surface area and porosity loss of the lime particles. The number of process cycles is important when considering cost of limestone feed; however, cost is not considered in this study.

As with MEA, CaO has a higher affinity for SO_2 than CO_2 . Loss of available CaO surface area due to the reaction with SO_2 in the flue gas was calculated at 1.4 wt % per cycle (Appendix C3). Additional limestone to make up for this lost CaO was not included in the process simulation because it does not have a significant impact on the energy performance analysis. Also, the lime will not react with water at the temperature and partial pressure of water vapor in the carbonator.⁴²

Lime solids were transported to the carbonator using a portion of the boiler flue gas (Appendix C3), while limestone solids were transported to the calciner using the CO_2 /oxygen/NG gas product stream (Appendix C4). Additional fans were used for these gas streams to make up for pressure loss due to transport. A cyclone was assumed to be appropriate for separating the limestone product from the gas stream.

Calciner:

The limestone generated in the carbonator is transported to the calciner for lime regeneration. As with the carbonator, FBR was selected as the most practical reactor type for the calciner. Several alternatives were considered and dismissed for the type of reactor: a rotary kiln would have lower throughput and less efficient mixing for mass and heat transfer; entrained bed and fixed bed are limited by flow capacity; a fixed bed would also require periodic swing between adsorption and desorption involving a significant temperature change and thus a long transition time; a heat carrier solid was also

considered, but was not selected due to fluidization requirements and the necessary segregation of heat carrier solid particles from lime particles.

The temperature in the calciner was maintained at 950 °C in order to convert CaCO_3 to CaO . Using the $\text{CaCO}_3/\text{CO}_2$ equilibrium curve in Appendix A1, this temperature is reasonable for an outlet CO_2 partial pressure of over 14.7 psia (1.0 atm). The desorption of CO_2 is an endothermic reaction. Therefore, the temperature in the calciner was held constant by combusting natural gas (NG) with oxygen (95% O_2 purity) in a combustion chamber (indicated by a box at the bottom of the “Carbonator” in Figure 10) immediately before the reactor. The combustion of natural gas provided direct heating in the calciner. Coal was considered as an alternative to natural gas, but was not selected because coal combustion would introduce impurities such as SO_2 , PM, and mercury into the pure CO_2 stream, requiring additional control systems. As in the carbonator, for gas-solid heat transfer the heat transfer efficiency between the combusted NG/oxygen stream and limestone is assumed to be 100%. A calculation for the pressure drop required in the calciner (Appendix C4), which provides fluidization of the particles, shows that 0.08 atm (1.18 psi) is reasonable. This pressure drop assumes a reactor height of 30 m.

As mentioned above, this study assumes an ideal cycle for adsorption/desorption by CaO , with zero CaCO_3 waste. However, in practice there would be some amount of CaCO_3 bleed from the system. Abanades used a bleed rate of 0.1 kmol CaCO_3 equivalent per kmol flue CO_2 .⁴⁰ Effects of boiler fly ash on the LSCC system, such as a decreased temperature in the calciner, were not considered in this study.

As shown in Figure 10, a portion of the CO_2 -rich stream from the calciner outlet is circulated to mix with the NG/oxygen combustion gases in the calciner. The purpose of this CO_2 -rich stream is to provide an oxygen-containing gas mixture comparable to an air-blown PC, which helps to ensure that the flame temperature remains within the calciner equipment temperature limit.¹⁴ The flow rate of this stream is dependent on the oxygen concentration set-point in the calciner, which for this study is 50 mol %.⁴⁴ A water/gas heat exchanger is required to cool the CO_2 - H_2O stream, and a condenser is required to remove H_2O before the CO_2 compressor. The heat exchanger is integrated with the plant steam cycle and produces additional high-pressure, high-temperature steam

for the plant at 540 °C and 2,415 psia (164 atm), based on a simulated amount of feed water (Section 2.2.3). This configuration for heat recovery is intended for energy calculation purposes, and the practical configuration of the heat exchanger may be different. Potential negative effects of contaminants on the CO₂ product stream pipeline, such as from nitrogen and oxygen, were not considered in this study.

Plant Steam Cycle:

The operating conditions of the plant's sub-critical steam cycle were modified from the base plant in order to accommodate integration with the LSCC process. Providing more available steam to the HP turbine to generate electricity – from the carbonator and heat exchanger after the calciner – was simulated in this study. Additional details and assumptions for the plant steam cycle are provided in Appendix C.

Preliminary Comparison of LSCC Process to MEA Process:

There are some notable differences between the LSCC process and MEA process: duct modification after the economizer and before the SCR; carrier material (CaO) for CO₂ capture; equipment for solids transport; condenser and heat exchanger duties; and the air separation unit addition. Table 6 is a qualitative, summary comparison of the major operating equipment and parameters for the LSCC and MEA processes.

Table 6 Qualitative comparison of LSCC and MEA processes

Variable	PC plant with MEA	PC plant with LSCC
<i>Boiler/duct</i>		
Boiler/duct modifications	None	Reconfiguration of ducting between economizer and SCR
<i>CO₂ capture unit</i>		
Reactor type	Packed bed absorption column	Circulating fluidized bed
CO ₂ capture material	Monoethanolamine (liquid)	Lime (solid)
<i>CO₂ desorption</i>		
Reactor type	Stripper with reboiler	Circulating fluidized bed
Regeneration heat method	Low-pressure steam	NG/O ₂ combustion
<i>CO₂ preparation</i>		
CO ₂ /H ₂ O condenser	Small heat duty	Large heat duty
<i>Steam cycle</i>		
High-pressure steam effects	None	HP steam generated from: heat recovery at carbonator; heat recovery from CO ₂ -lean flue gas at heat exchanger after carbonator; heat recovery from CO ₂ /water vapor gas stream at heat exchanger after calciner
Low-pressure steam effects	Steam extraction from intermediate-pressure (IP) turbine for solvent regeneration	None
<i>Others</i>		
Transfer of sorbent material	Liquid pumps	Portion of process gas + fan
SO ₂ control equipment	FGD required	FGD not required

The effects of these differences on the LSCC process performance, and the overall plant performance, are examined in the Results and Discussion section (Section 3.3).

Regarding effects of the LSCC process on the standard flue gas cleaning processes (SCR, ESP and FGD), only a couple of modifications to the base plant are required. First, a wet FGD for SO₂ removal is not necessary because the LSCC process is assumed to

effectively remove SO₂ from the boiler flue gas. Second, a larger induced draft fan is necessary due to the pressure drop of the post-boiler flue gas through the carbonator.

2.2.3. Simulation of LSCC Process

Simulation of the LSCC process for CO₂ capture was performed as a “proof-of-concept” study, with the goal of estimating the best-case energy use of the process at a PC power plant. ChemCAD software package (version 6.1.0) was used for the simulation. The four main process sections for simulating the coal-fired power plant were: Boiler section; Steam Turbine section; Gas Cleaning section; and Compression/Preparation section for CO₂ sequestration. The 533 MW_e (gross) base PC power plant, as well as the power plant equipped with the MEA process for CO₂ capture (hereafter referred to as PC-MEA), were referred to those reported by Chen *et al* which were based on a U.S. DOE PC plant.^{14,37} The base PC plant includes values for operating conditions in the four sections of the plant. The configuration and operating conditions of the base plant were modified to accommodate inclusion of the LSCC process (hereafter referred to as PC-LSCC). Table 7 gives the main parameter values used in the ChemCAD simulations for the PC-MEA and PC-LSCC plants. Energy usage for the different unit operations at the plants with LSCC and MEA were assumed to be the same as the DOE base plant, estimated from the DOE base plant, or found by direct simulation. Supporting information for the values in Table 7, as well as parameter values for the base PC plant, is provided in Appendices B and C.

Table 7 Main operating and performance parameters for MEA and LSCC processes

Parameter ^a	PC plant with MEA ¹⁴	PC plant with LSCC
<i>Steam generation</i>		
Gross output, MW _e	533	533
Hot reheat steam, lb/hr (kg/hr)	3,022,125 (1,370,578)	3,027,000 (1,372,789)
Superheat steam, lb/hr (kg/hr)	3,422,824 (1,552,301)	3,415,000 (1,548,753)
Additional steam to HP turbine from MEA/LSCC process, lb/hr (kg/hr)	0	2,976,000 (1,349,891)
Heat duty of cooling tower, mmBtu/hr (MJ/hr)	1,104 (1,164,720)	2,918 (3,078,800)
<i>Absorber/Carbonator</i>		
Inlet flue gas volume, lb/hr (kg/hr)	4,472,743 (2,028,455)	4,294,504 (1,947,621)
Gas inlet temperature, °C	54	340
SO ₂ inlet, lb/hr (kg/hr)	823 (373)	16,473 (7,471)
Active solvent or sorbent concentration, wt %	30	100
Active solvent or sorbent/CO ₂ molar ratio	5.0	1.0
Operating temperature, °C	40-57	650
Pressure drop, psi (atm)	3.00 (0.20)	1.47 (0.10)
CO ₂ captured, lb/hr (kg/hr) ^b	834,748 (378,571)	834,857 (378,620)
<i>Stripper/Calciner</i>		
Solvent or sorbent make-up, lb/hr (kg/hr)	2,794 (1,267)	0
LP steam consumption, lb/hr (kg/hr)	1,380,364 (626,015)	N/A ^c
Oxygen set point, mol %	N/A ^c	50
Oxygen flow rate (95% purity), lb/hr (kg/hr)	N/A ^c	353,700 (160,436)
Natural gas flow rate, lb/hr (kg/hr)	N/A ^c	83,500 (37,875)
Percent of flue gas product recycled	N/A ^c	16
Operating temperature, °C	95-116	950
Pressure drop, psi (atm)	3.00 (0.20)	1.18 (0.08)
^a Parameter definitions: Additional steam to HP turbine from MEA/LSCC process = HP steam generated from: heat recovery at adsorber; heat recovery from CO ₂ -lean flue gas at heat exchanger after carbonator; heat recovery from CO ₂ /water vapor gas stream at heat exchanger after calciner Gross output = electrical output from the turbine/generator set Heat duty of cooling tower = heat removed due to cooling and condensing steam from the steam turbines Hot reheat steam = HP steam from reheater in boiler LP steam consumption = LP steam required to provide the regeneration heat at the stripper for the MEA case		

Table 7 (cont.)

Percent of flue gas product recycled = the portion of the CO₂-rich stream from the calciner outlet recirculated back to the calciner

Solvent or sorbent make-up = fresh feed of solvent or sorbent into the CO₂ capture process

Superheat steam = HP steam from superheater in boiler

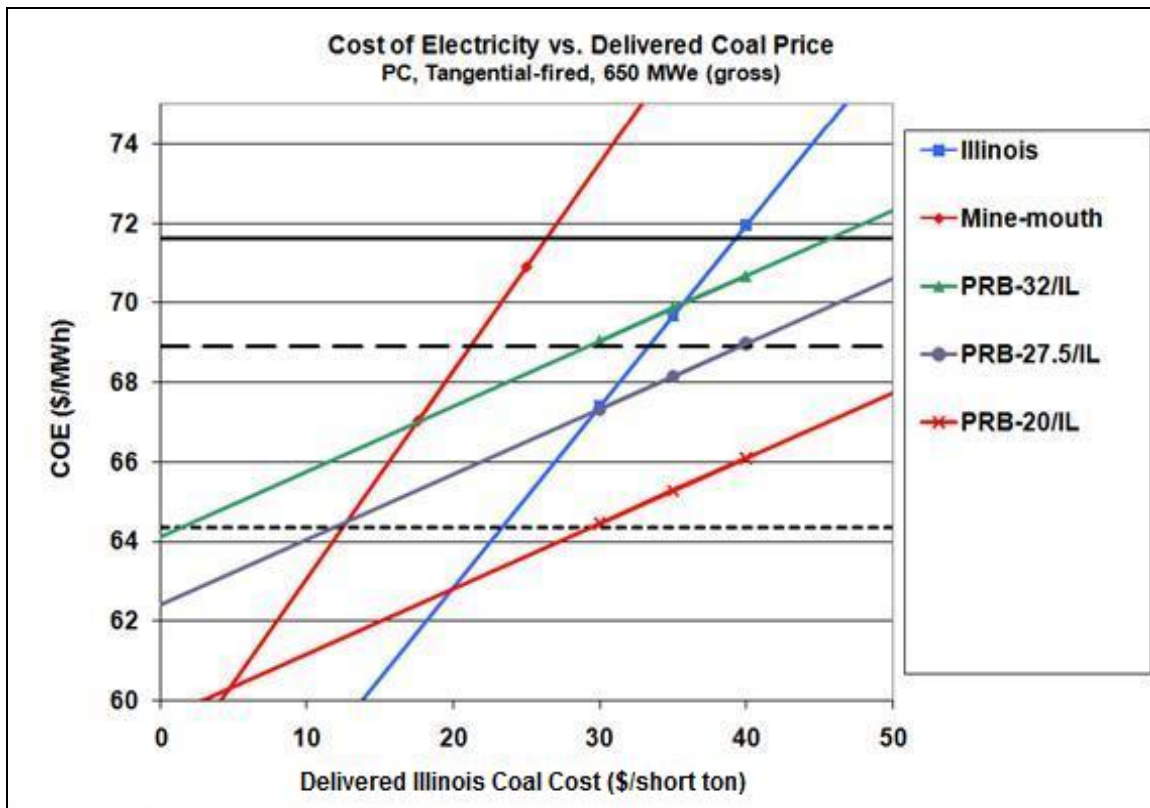
^b Difference in CO₂ captured between MEA and LSCC cases is due to round-off error

^c N/A = Not Applicable

3 RESULTS AND DISCUSSION

3.1 Scenario 1 - SO₂, NO_x, PM and Mercury Control with no CO₂ Capture

The case scenarios in Table 5 were evaluated using IECM. Figures 11 and 12 show COE values for the 650 MW_e (gross) and 175 MW_e (gross) cases. These figures represent COE values for a power plant along the vertical axis versus delivered Illinois coal costs along the horizontal axis when firing different types of fuel (Illinois, PRB or 70/30 PRB/IL blend coals), for fixed delivered PRB coal costs of \$20.0, \$27.5 and \$32.0/short ton (\$22.0, \$30.3 and \$35.3/metric ton). The COE values for a power plant firing 100% PRB coal are not dependent on the delivered Illinois coal cost, so those costs are represented by a constant COE value for each of the delivered PRB coal costs (\$20.0, \$27.5 and \$32.0/short ton, or \$22.0, \$30.3 and \$35.3/metric ton). The breakeven COE value for a particular coal type can be determined by finding the delivered Illinois coal cost below which the corresponding plant COE value is less than the plant COE value of an alternative coal type and price. For example, in Figure 11 for 100% Illinois coal, a delivered Illinois coal cost less than \$33.0/short ton (\$36.4/metric ton) corresponds to a COE value less than the COE value for 100% PRB coal at \$27.5/short ton (\$30.3/metric ton). Thus, the breakeven COE value for 100% Illinois coal against a fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton) occurs at a delivered Illinois coal cost of \$33.0/short ton (\$36.4/metric ton). Table 8 shows the breakeven delivered costs of each coal type (100% Illinois, Illinois mine-mouth, and 70/30 PRB/IL coal blend), for a fixed PRB coal delivered cost of \$27.5/short ton (\$30.3/metric ton). Capital and O&M costs were also estimated (Appendix D).



Notes:

COE = Cost of Electricity

Illinois = 100% Illinois coal

Mine-mouth = Illinois mine-mouth coal

PRB-32/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$32.0/short ton (\$35.3/metric ton)

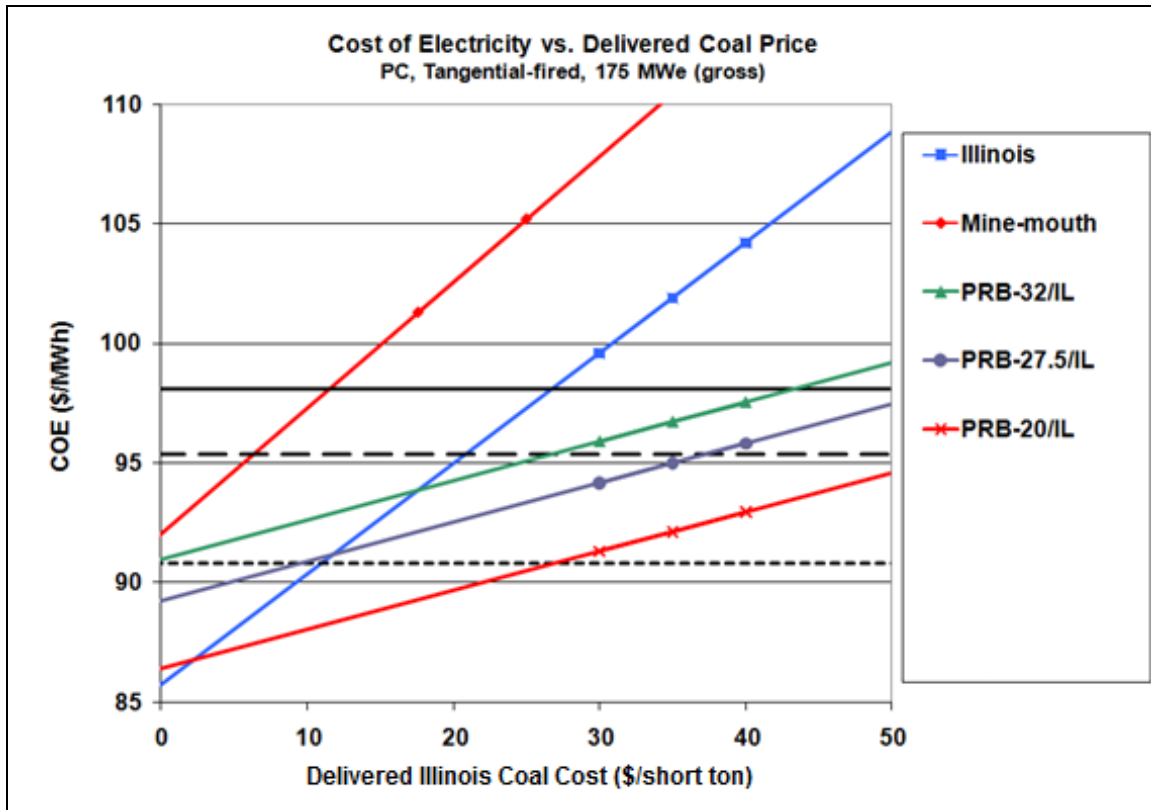
PRB-27.5/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$20.0/short ton (\$22.0/metric ton)

The three horizontal lines – distinct from the vertical axis marks – represent COE values at different delivered PRB coal costs:

<hr style="border: none; border-top: 2px solid black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$32.0/short ton (\$35.3/metric ton)
<hr style="border: none; border-top: 2px dashed black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$27.5/short ton (\$30.3/metric ton)
<hr style="border: none; border-top: 2px dashed black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$20.0/short ton (\$22.0/metric ton)

Figure 11 COE versus delivered Illinois coal cost for 650 MWe (gross) plant



Notes:

COE = Cost of Electricity

Illinois = 100% Illinois coal

Mine-mouth = Illinois mine-mouth coal

PRB-32/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$32.0/short ton (\$35.3/metric ton)

PRB-27.5/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$20.0/short ton (\$22.0/metric ton)

The three horizontal lines – distinct from the vertical axis marks – represent COE values at different delivered PRB coal costs:

<hr style="border: none; border-top: 2px solid black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$32.0/short ton (\$35.3/metric ton)
<hr style="border: none; border-top: 2px dashed black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$27.5/short ton (\$30.3/metric ton)
<hr style="border: none; border-top: 2px dotted black; width: 100px; margin-bottom: 5px;"/>	COE value for 100% PRB coal at \$20.0/short ton (\$22.0/metric ton)

Figure 12 COE versus delivered Illinois coal cost for 175 MWe (gross) plant

Table 8 Breakeven delivered Illinois coal costs against a fixed PRB cost of \$27.5/short ton (\$30.3/metric ton) (PC-tangential fired boiler)

Coal type	Breakeven delivered Illinois cost, \$/short ton (\$/metric ton)		2007 market cost of Illinois coal, \$/short ton (\$/metric ton)
	650 MW _e (gross) ^a	175 MW _e (gross) ^a	
PRB-27.5/IL blend ^b	39.5 (43.5)	36.0 (39.7)	35.0 (38.6)
Mine-mouth IL	21.0 (23.1)	6.5 (7.2)	17.6 (19.4)
100% Illinois	33.0 (36.4)	21.0 (23.1)	35.0 (38.6)
^a “Gross” indicates <i>gross electrical output</i> , which means the total electrical output from an electric generating unit (EGU) before making any deductions for energy output used in any way related to the production of energy (for an EGU generating only electricity, the gross electrical output is the output from the turbine/generator set) ^b PRB-27.5/IL blend = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)			

For 100% PRB coal at \$27.5/short ton (\$30.3/metric ton), the breakeven coal costs of 100% Illinois coal are less than the 2007 Illinois coal market cost for the 650 MW_e case. This indicates that the 100% Illinois coal case does not compete with the 100% PRB coal case. However, for the 70/30 PRB/IL coal blend, the breakeven coal costs of Illinois coal are higher than the 2007 Illinois coal market costs, making Illinois coal more competitive with 100% PRB. As the PRB coal cost increases, the breakeven delivered Illinois coal, Illinois mine-mouth coal, and 70/30 PRB/IL coal blend costs increase and those coals become more competitive with the 100% PRB coal case. For example, at a delivered PRB coal cost of \$32.0/short ton (\$35.3/metric ton) at a 650 MW_e plant, the breakeven delivered Illinois coal cost is \$39.5/short ton (\$43.5/metric ton) for the 100% Illinois coal case. Similar trends can be observed for the 175 MW_e case. For both plant sizes, the 70/30 PRB/IL coal blend is more competitive with PRB coal than 100% Illinois coal, at all the selected PRB delivered costs (\$20.0, \$27.5 and \$32.0/short ton, or \$22.0, \$30.3 and \$35.3/metric ton).

For the 650 MW_e case, at constant PRB coal costs of \$27.5/short ton (\$30.3/metric ton) and \$32.0/short ton (\$35.3/metric ton), the addition of Illinois coal to PRB coal to provide a 70/30 PRB/IL coal blend gives comparable (within \$1.8/short ton, or \$2.0/metric ton) or less COE values compared to 100% PRB coal, for the selected range of Illinois coal costs (\$30.0, \$35.0 and \$40.0/short ton, or \$33.1, \$38.6 and \$44.1/metric

ton). This shows that the 70/30 PRB/IL coal blend can be competitive with 100% PRB coal for the 650 MW_e case. Similar trends can be observed for the 175 MW_e case. However, breakeven Illinois coal costs for the 100% Illinois and Illinois mine-mouth cases (\$21.0/short ton and \$6.5/short ton (\$23.1/metric ton and \$7.2/metric ton), respectively) at a constant PRB coal cost of \$27.5/short ton (\$30.3/metric ton) are distinctively different from those at the 650 MW_e plant (\$33.0/short ton and \$21.0/short ton, or \$36.4/metric ton and \$23.1/metric ton). A likely reason for this is that the SO₂ and PM control costs for these two cases have a higher contribution to COE at the 175 MW_e plant than at the 650 MW_e plant, whereas the control costs for the 70/30 PRB/IL coal blend have essentially the same contribution at both plant sizes (see Figures 13 and 14 below). An Illinois mine-mouth plant can be competitive with the PRB and 70/30 PRB/IL coal blend only at the 650 MW_e plant size. For this plant size, a low-cost Illinois mine coal (\$17.6/short ton, or \$19.4/metric ton) yields lower COE values than 100% PRB and 70/30 PRB/IL blend coals, except for some cases with the lowest PRB coal cost assumed (\$20.0/short ton, or \$22.0/metric ton). At a high Illinois mine-mouth coal cost (\$25.0/short ton, or \$27.6/metric ton), the 650 MW_e plant has comparable or lower COE values than the following cases: 100% PRB coal case with the highest cost at \$32.0/short ton (\$35.3/metric ton); and the two 70/30 PRB/IL coal blend cases with the costs of Illinois coal at \$35.0/short ton and \$40.0/short ton (\$38.6/metric ton and \$44.1/metric ton) and the cost of PRB at \$32.0/short ton (\$35.3/metric ton).

The breakdown of costs (Figures 13 and 14) shows that plant COE is most sensitive to fuel cost and SO₂ control cost – due to a high percentage contribution of “fuel” and “SO₂” in the total cost – and PM control cost to a lesser extent. This high sensitivity is due primarily to the significant variation in both the coal cost (\$17.6 - \$40.0/short ton, or \$19.4 - \$44.1/metric ton) and coal sulfur content (0.300 - 4.000 wt %) among the different coals. Plant COE is not sensitive to NO_x control cost, and mercury control does not show significant influence on the plant COE. Also, a 70/30 PRB/IL coal blend has a slightly lower percentage contribution of control cost to COE than 100% Illinois coal because it employs SDA+FF rather than wet FGD+ESP.

The most cost-effective scenario for a 650 MW_e PC plant – at 2007 market coal costs of \$27.5/short ton (\$30.3/metric ton) for PRB coal, \$35.0/short ton (\$38.6/metric ton) for

Illinois coal, and \$17.6/short ton (\$19.4/metric ton) for Illinois mine-mouth coal – is the Illinois mine-mouth coal (\$67.0/MWh). The Illinois mine-mouth coal is most cost-effective for the 650 MW_e case due primarily to the lower contribution of “fuel” to plant COE compared to the “fuel” contributions of the other coal types (Figure 13). This lower “fuel” contribution helps compensate for the higher contribution of SO₂ control to plant COE for the Illinois mine-mouth coal compared to the SO₂ control contributions of the other coal types. The most cost-effective scenario for a 175 MW_e PC plant – at 2007 market coal costs – is the 70/30 PRB/IL coal blend (\$95.0/MWh). The 70/30 PRB/IL coal blend replaces the Illinois mine-mouth as most cost-effective when the plant size is reduced to 175 MW_e – even though the Illinois mine-mouth coal still has a lower contribution of “fuel” to plant COE compared to the other coal types – for the following reason: the SO₂ and PM control costs for the Illinois mine-mouth case have a higher contribution to plant COE at the 175 MW_e plant than at the 650 MW_e plant, whereas these control costs for the 70/30 PRB/IL coal blend have essentially the same contribution to plant COE at both plant sizes (Figure 14).

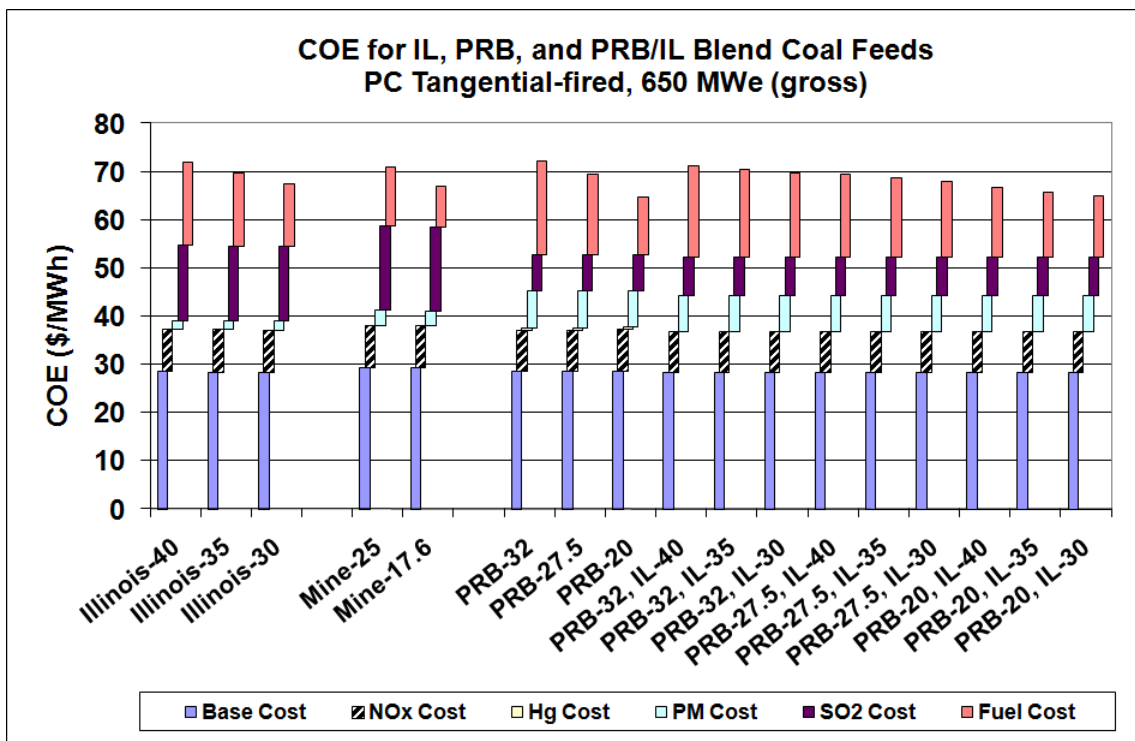


Figure 13 Breakdown costs for 650 MW_e (gross) PC plant

Notes:

COE = Cost of Electricity

Illinois-40 = 100% Illinois coal at cost of \$40.0/short ton (\$44.1/metric ton)

Illinois-35 = 100% Illinois coal at cost of \$35.0/short ton (\$38.6/metric ton)

Illinois-30 = 100% Illinois coal at cost of \$30.0/short ton (\$33.1/metric ton)

Mine-25 = Illinois mine-mouth coal at cost of \$25.0/short ton (\$27.6/metric ton)

Mine-17.6 = Illinois mine-mouth coal at cost of \$17.6/short ton (\$19.4/metric ton)

PRB-32 = 100% PRB coal at cost of \$32.0/short ton (\$35.3/metric ton)

PRB-27.5 = 100% PRB coal at cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20 = 100% PRB coal at cost of \$20.0/short ton (\$22.0/metric ton)

PRB-32, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-32, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-32, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

PRB-27.5, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-27.5, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

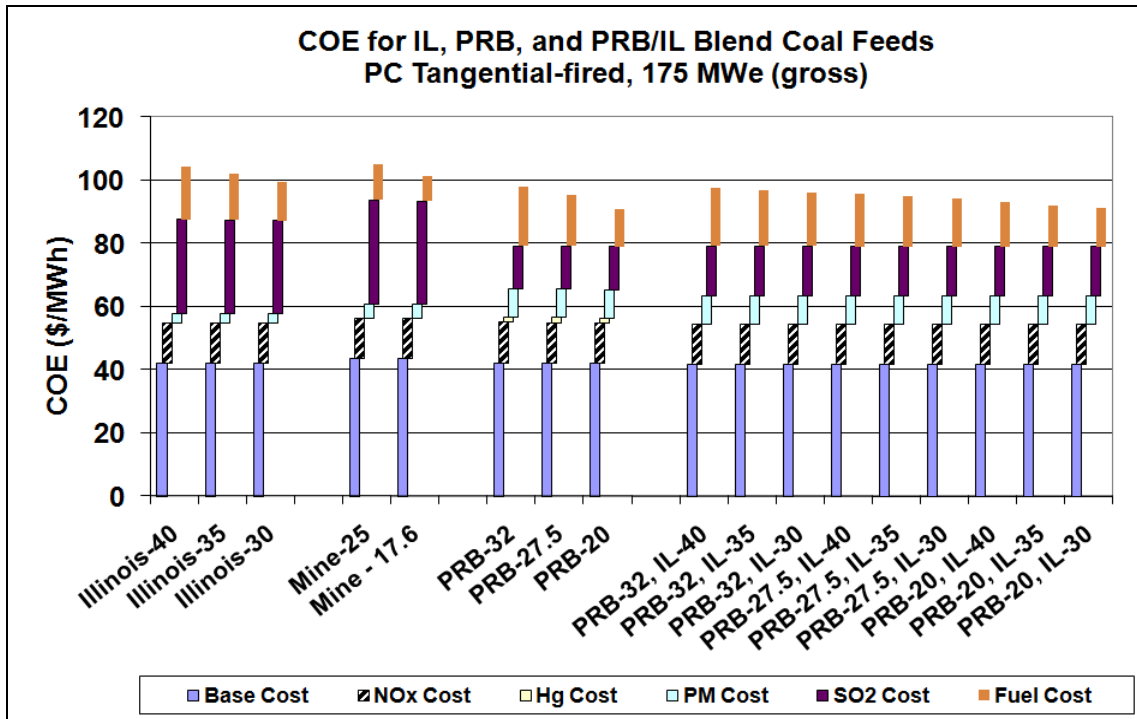
PRB-27.5, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

PRB-20, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-20, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-20, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

Figure 13 (cont.)



Notes:

COE = Cost of electricity

Illinois-40 = 100% Illinois coal at cost of \$40.0/short ton (\$44.1/metric ton)

Illinois-35 = 100% Illinois coal at cost of \$35.0/short ton (\$38.6/metric ton)

Illinois-30 = 100% Illinois coal at cost of \$30.0/short ton (\$33.1/metric ton)

Mine-25 = Illinois mine-mouth coal at cost of \$25.0/short ton (\$27.6/metric ton)

Mine-17.6 = Illinois mine-mouth coal at cost of \$17.6/short ton (\$19.4/metric ton)

PRB-32 = 100% PRB coal at cost of \$32.0/short ton (\$35.3/metric ton)

PRB-27.5 = 100% PRB coal at cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20 = 100% PRB coal at cost of \$20.0/short ton (\$22.0/metric ton)

PRB-32, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-32, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-32, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

PRB-27.5, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-27.5, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

Figure 14 Breakdown costs for 175 MWe (gross) PC plant

PRB-27.5, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.0/metric ton)

PRB-20, IL-40 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

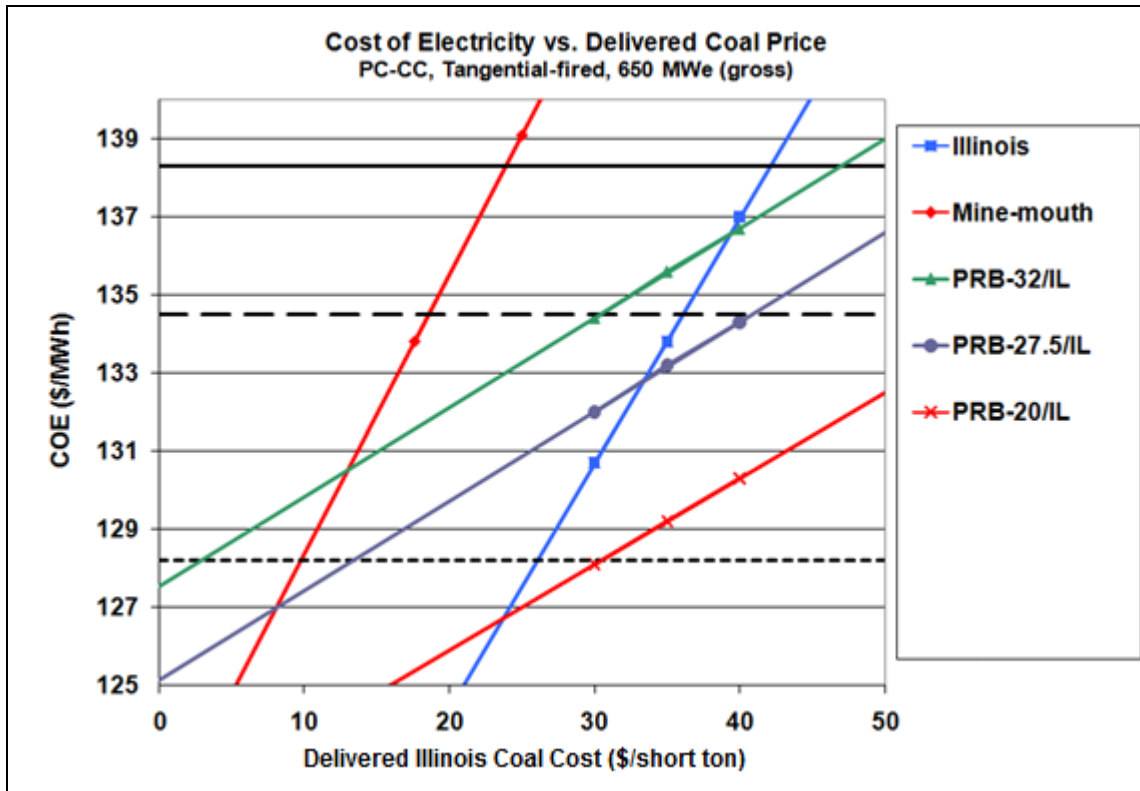
PRB-20, IL-35 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-20, IL-30 = 70/30 blend of PRB/IL coal by weight at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

Figure 14 (cont.)

3.2 Scenario 2 - SO₂, NO_x, PM and Mercury Control and 90% CO₂ Capture with MEA

Figures 15 and 16 show new COE values for the previous PC power plant scenarios, with the addition of a CO₂ capture system (CC). Table 9 shows the breakeven delivered costs of each coal type, for a fixed PRB delivered cost of \$27.5/short ton (\$30.3/metric ton). Tables of environmental as well as economic impacts of CO₂ capture on the 650 MW_e (gross) and 175 MW_e (gross) PC plant cases are in Appendix D.



Notes:

COE = Cost of electricity

Illinois = 100% Illinois coal

Mine-mouth = Illinois mine-mouth coal

PRB-32/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$32.0/short ton (\$35.3/metric ton)

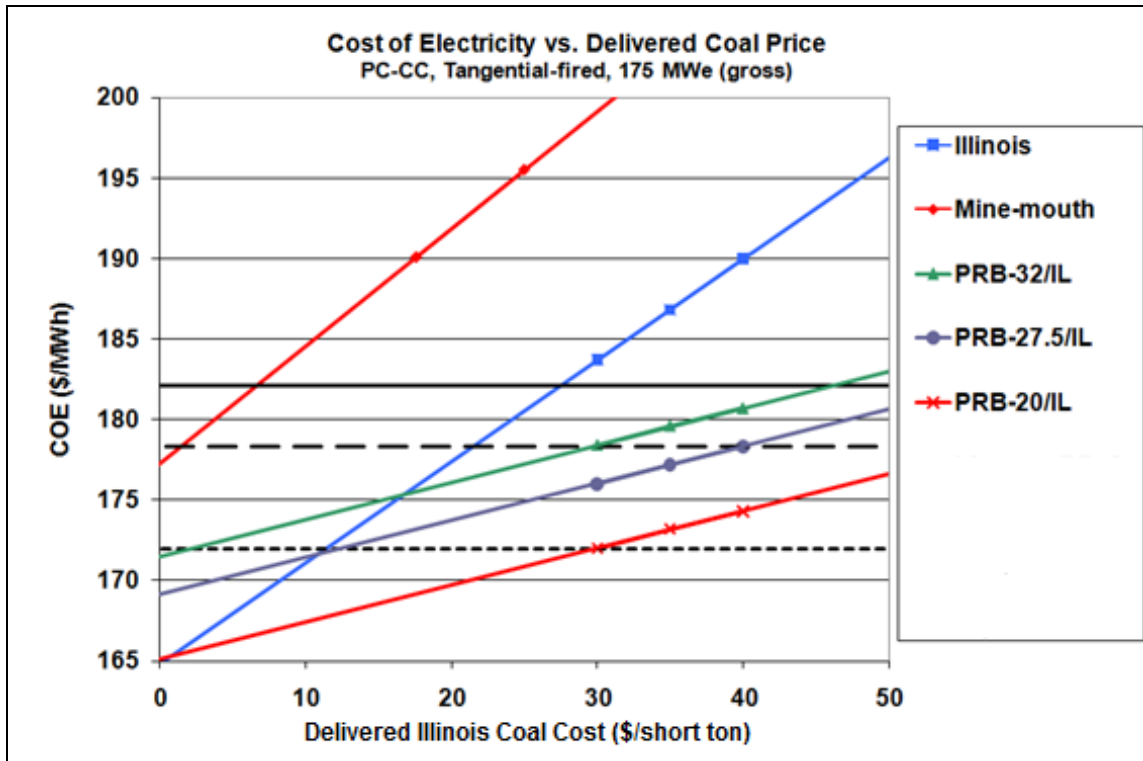
PRB-27.5/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$20.0/short ton (\$22.0/metric ton)

The three horizontal lines – distinct from the vertical axis marks – represent COE values at different delivered PRB coal costs:

—————	COE value for 100% PRB coal at \$32.0/short ton (\$35.3/metric ton)
- - - - -	COE value for 100% PRB coal at \$27.5/short ton (\$30.3/metric ton)
- - - - -	COE value for 100% PRB coal at \$20.0/short ton (\$22.0/metric ton)

Figure 15 COE versus delivered Illinois coal cost for 650 MWe (gross) PC-CC plant



Notes:

COE = Cost of electricity

Illinois = 100% Illinois coal

Mine-mouth = Illinois mine-mouth coal

PRB-32/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$32.0/short ton (\$35.3/metric ton)

PRB-27.5/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20/IL = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$20.0/short ton (\$22.0/metric ton)

The three horizontal lines – distinct from the vertical axis marks – represent COE values at different delivered PRB coal costs:

—————	COE value for 100% PRB coal at \$32.0/short ton (\$35.3/metric ton)
- - - - -	COE value for 100% PRB coal at \$27.5/short ton (\$30.3/metric ton)
- - - - -	COE value for 100% PRB coal at \$20.0/short ton (\$22.0/metric ton)

Figure 16 COE versus delivered Illinois coal cost for 175 MWe (gross) PC-CC plant

Table 9 Breakeven delivered Illinois coal costs against a fixed PRB cost of \$27.5/short ton (\$30.3/metric ton) (PC-tangential fired boiler with CC)

Coal type	Breakeven delivered Illinois cost, \$/short ton (\$/metric ton)		2007 market cost of Illinois coal, \$/short ton (\$/metric ton)
	650 MW _e (gross) ^a	175 MW _e (gross) ^a	
PRB-27.5/IL blend ^b	41.0 (45.2)	39.5 (43.5)	35.0 (38.6)
Mine-mouth IL	18.5 (20.4)	1.5 (1.7)	17.6 (19.4)
100% Illinois	36.0 (39.7)	21.5 (23.7)	35.0 (38.6)
^a “Gross” indicates <i>gross electrical output</i> , which means the total electrical output from an electric generating unit (EGU) before making any deductions for energy output used in any way related to the production of energy (for an EGU generating only electricity, the gross electrical output is the output from the turbine/generator set) ^b PRB-27.5/IL blend = 70/30 blend of PRB/IL coal by weight at fixed PRB coal cost of \$27.5/short ton (\$30.3/metric ton)			

As shown in Appendix D, the SO₂ emissions from the plant dropped sharply (> 99.5% on a short tons/hour and metric tons/hour basis) with the addition of CO₂ capture, due to the strong affinity of the amine solvent for SO₂. Emissions of NO_x dropped slightly (2%) with the addition of CO₂ capture, while PM emissions dropped 50% and mercury emissions remained the same, on a short tons/hour and metric tons/hour basis. The higher total CO₂ emissions (additional 4%) when burning PRB coal compared to Illinois coal is due to the lower higher heating value (HHV) and higher moisture content of the PRB coal. The higher CO₂ emissions, combined with a higher flue gas volume (i.e., 5%) compared to Illinois coal, results in a higher control cost when burning PRB coal.

Addition of CC to the PC scenarios adds \$63/MWh – \$68/MWh to the COE values for the 650 MW_e case, and \$72/MWh – \$90/MWh to the COE values for the 175 MW_e case. The 70/30 PRB/IL coal blend has the least total cost for both fixed O&M and variable O&M, while 100% Illinois coal has the least capital cost (Appendix D). For 100% PRB coal at a delivered cost of \$27.5/short ton (\$30.3/metric ton), the breakeven cost of 100% Illinois coal for the 650 MW_e plant (\$36.0/short ton, or \$39.7/metric ton) increased to above the 2007 Illinois coal market cost (\$35.0/short ton, or \$38.6/metric ton). This shows that 100% Illinois coal can compete well with PRB in this case for a PC-CC plant, whereas it could not compete with PRB for a PC plant without CO₂ capture. For the 70/30 PRB/IL coal blend, there is again an increase in the breakeven costs of Illinois coal

– where it is still higher than the 2007 Illinois market cost – and thus the 70/30 PRB/IL coal blend remains competitive with PRB, as with the PC plant without CO₂ capture.

As the PRB coal cost increases, the breakeven delivered Illinois coal, Illinois mine-mouth coal, and 70/30 PRB/IL coal blend costs increase and those coals become more competitive with PRB coal. As with the 650 MW_e PC plant, at constant PRB coal costs of \$27.5/short ton and \$32.0/short ton (\$30.3/metric ton and \$35.3/metric ton), the addition of Illinois coal to PRB coal in a 70/30 PRB/IL coal blend gives comparable (within \$2.2/short ton, or \$2.4/metric ton) or less COE values compared to 100% PRB coal, for the selected range of Illinois coal costs (\$30.0, \$35.0 and \$40.0/short ton, or \$33.1, \$38.6 and \$44.1/metric ton). This shows that the 70/30 PRB/IL coal blend can be competitive with 100% PRB coal for the PC-CC plant. As with the PC plant without CO₂ capture, trends similar to the 650 MW_e case with CO₂ capture can be observed for the 175 MW_e case with CO₂ capture. However, the breakeven Illinois coal costs for the 100% Illinois and Illinois mine-mouth cases (\$21.5/short ton and \$1.5/short ton (\$23.7/metric ton and \$1.7/metric ton), respectively) are again distinctively different from those at the 650 MW_e plant (\$36.0/short ton and \$21.5/short ton, or \$39.7/metric ton and \$23.7/metric ton). The likely reason for this is the same as for the PC plants without CO₂ capture: the SO₂ and PM control costs for these two cases have a higher contribution to COE at the 175 MW_e plant compared to the 650 MW_e plant, whereas the control costs for the 70/30 PRB/IL coal blend have essentially the same contribution to COE at both plant sizes (Figures 17 and 18).

An Illinois mine-mouth plant (at \$17.6/short ton (\$19.4/metric ton) coal cost) can be competitive with 100% PRB coal and 70/30 PRB/IL coal blend, but only for the 650 MW_e plant tested here. For this plant size, a low-cost mine-mouth coal (\$17.6/short ton, or \$19.4/metric ton) yields lower COE values than the following cases: 70/30 PRB/IL coal blend at \$32.0/short ton (\$35.3/metric ton) PRB coal cost; 70/30 PRB/IL coal blend at \$27.5/short ton (\$30.3/metric ton) PRB coal cost and \$40.0/short ton (\$44.1/metric ton) Illinois coal cost; and 100% Illinois coal at \$35.0 and \$40.0/short ton (\$38.6 and \$44.1/metric ton) coal costs. At high Illinois mine-mouth coal cost (\$25.0/short ton, or \$27.6/metric ton), this 650 MW_e plant has higher COE values than all other coal types and costs.

The best case scenario for PC plants seeking to add CO₂ capture, at 2007 market coal costs (\$35.0/short ton (\$38.5/metric ton) for Illinois coal, \$27.5/short ton (\$30.3/metric ton) for PRB coal, and \$17.6/short ton (\$19.4/metric ton) for Illinois mine-mouth coal), is the 70/30 PRB/IL coal blend for both a 650 MW_e plant (\$133.1/MWh) and a 175 MW_e plant (\$177.0/MWh). The CO₂ control cost contribution to plant COE for the 70/30 PRB/IL coal blend is higher than those of the 100% Illinois and Illinois mine-mouth coals, due to the higher CO₂ emissions inherent to burning PRB coal. However, the higher SO₂ control cost contributions to COE for the 100% Illinois and Illinois mine-mouth coals, compared to the 70/30 PRB/IL coal blend, counter the lower CO₂ control cost contributions for these coals sufficiently to make the 70/30 PRB/IL coal blend most cost-effective for the 650 MW_e and 175 MW_e cases (Figures 17 and 18).

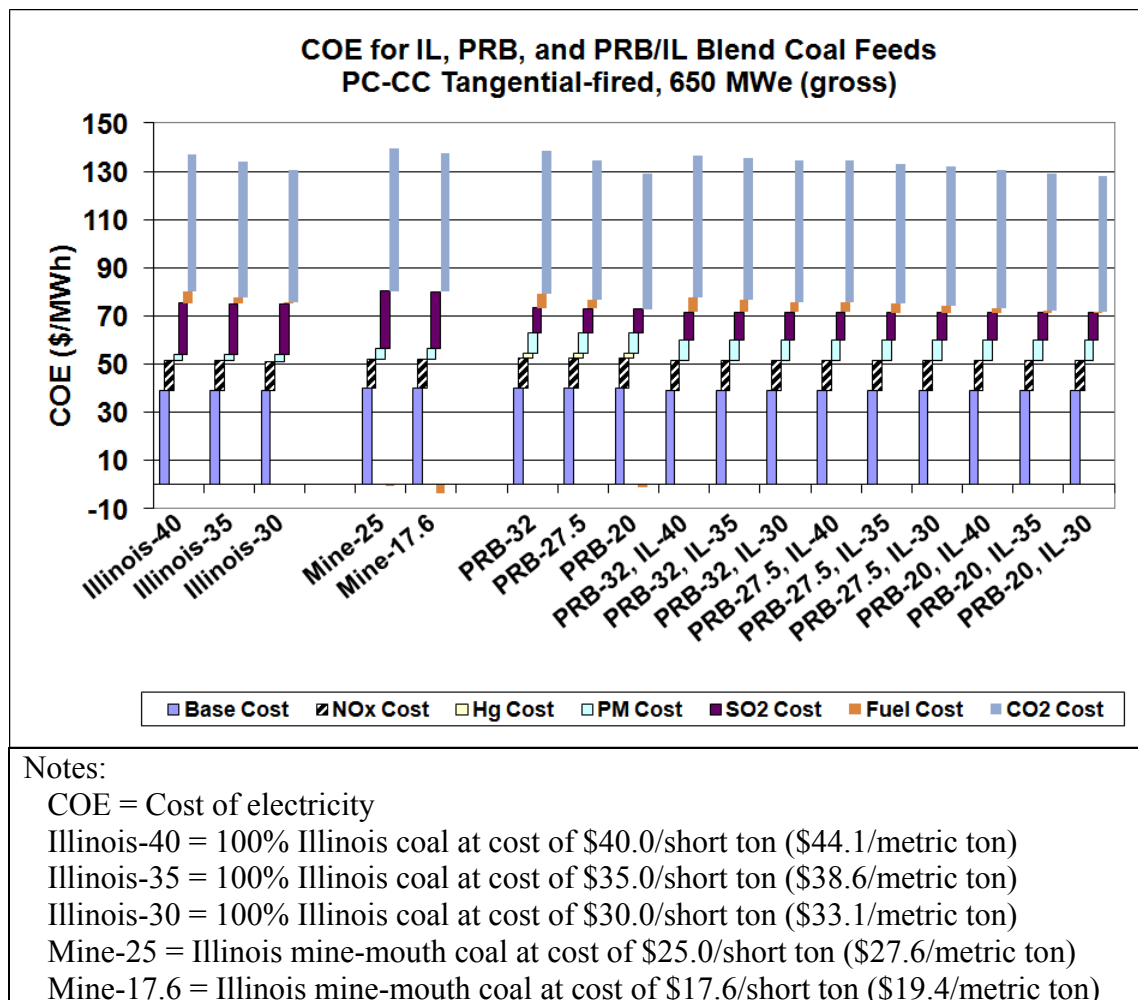
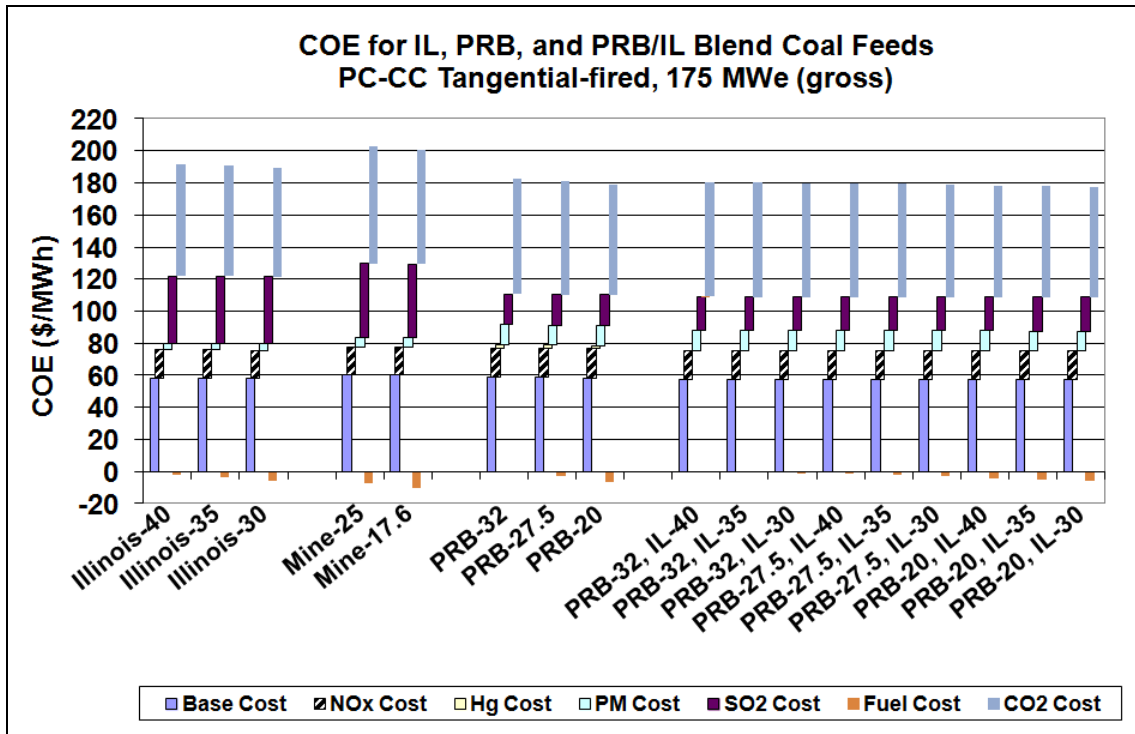


Figure 17 Breakdown costs for 650 MW_e (gross) PC-CC plant

PRB-32 = 100% PRB coal at cost of \$32.0/short ton (\$35.3/metric ton)
 PRB-27.5 = 100% PRB coal at cost of \$27.5/short ton (\$30.3/metric ton)
 PRB-20 = 100% PRB coal at cost of \$20.0/short ton (\$22.0/metric ton)
 PRB-32, IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)
 PRB-32, IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)
 PRB-32, IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)
 PRB-27.5, IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)
 PRB-27.5, IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)
 PRB-27.5, IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)
 PRB-20, IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)
 PRB-20, IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)
 PRB-20, IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)
 The IECM charges each abatement technology for the internal use of electricity and treats the charge as a credit within the “Boiler O&M” cost component of the model. Thus the contribution of Fuel Cost – which is based on the “Boiler O&M” component – to COE is very low or negative when CO₂ capture technology is implemented

Figure 17 (cont.)



Notes:

COE = Cost of electricity

Illinois-40 = 100% Illinois coal at cost of \$40.0/short ton (\$44.1/metric ton)

Illinois-35 = 100% Illinois coal at cost of \$35.0/short ton (\$38.6/metric ton)

Illinois-30 = 100% Illinois coal at cost of \$30.0/short ton (\$33.1/metric ton)

Mine-25 = Illinois mine-mouth coal at cost of \$25.0/short ton (\$27.6/metric ton)

Mine-17.6 = Illinois mine-mouth coal at cost of \$17.6/short ton (\$19.4/metric ton)

PRB-32 = 100% PRB coal at cost of \$32.0/short ton (\$35.3/metric ton)

PRB-27.5 = 100% PRB coal at cost of \$27.5/short ton (\$30.3/metric ton)

PRB-20 = 100% PRB coal at cost of \$20.0/short ton (\$22.0/metric ton)

PRB-32/IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-32/IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-32/IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$32.0/short ton (\$35.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

PRB-27.5/IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

PRB-27.5/IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)

PRB-27.5/IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$27.5/short ton (\$30.3/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)

PRB-20/IL-40 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$40.0/short ton (\$44.1/metric ton)

Figure 18 Breakdown costs for 175 MWe (gross) PC-CC plant

PRB-20/IL-35 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$35.0/short ton (\$38.6/metric ton)
 PRB-20/IL-30 = 70/30 blend of PRB/IL coal at PRB coal cost of \$20.0/short ton (\$22.0/metric ton) and Illinois coal cost of \$30.0/short ton (\$33.1/metric ton)
 The IECM charges each abatement technology for the internal use of electricity and treats the charge as a credit within the “Boiler O&M” cost component of the model. Thus the contribution of Fuel Cost – which is based on the “Boiler O&M” component – to COE is very low or negative when CO₂ capture technology is implemented

Figure 18 (cont.)

3.3 Assessment of Limestone Carbonation-Calcination Process

The LSCC process for CO₂ capture from the boiler flue gas, along with the boiler, steam turbines, and gas cleaning systems, were evaluated using ChemCAD (version 6.1.0). The inputs from Table 7, along with additional inputs found in Appendix C, were used in the ChemCAD program to obtain values for process energy usage. Table 10 shows the results of auxiliary power use and overall process performance for a 533 MW_e (gross) plant with LSCC. The case with no CO₂ capture, and the case with CO₂ capture using the MEA process, is also included in the table for comparison. A summary comparison of process conditions and environmental impacts for the LSCC and MEA processes is presented in this section as well. Detailed results of the mass/energy balance simulations for the LSCC process are available in Appendix E.

Table 10 Auxiliary power use and overall process performance for select cases

Auxiliary load, kW _e ^{a,b}	PC-LSCC	PC-MEA ¹⁴	Base plant with no CO ₂ capture ¹⁴
Coal handling	233	233	233
Coal pulverizer	2,017	2,017	2,017
Primary air fans	1,199	1,199	1,199
Forced draft fans	1,142	1,142	1,142
Induced draft fans	4,182 ^d	4,921	4,921
Seal air blowers	45	45	45
Steam turbine auxiliaries	884	884	884
Steam condensate pumps	895	895	895
Main feed pump	20,706 ^d	10,938 ^c	10,938 ^c
Circulating water pumps	8,634	2,124 ^e	4,187
Cooling tower fans	4,881	1,201 ^e	2,367
Ash handling	1,658	1,658	1,658
Miscellaneous	2,411	2,411	2,411
Transformer loss	1,215	1,215	1,215
LSCC or MEA:			
<i>Induced draft fans</i>	11,90 ^d	13,098 ^d	N/A ^f
<i>Sorbent or solvent transportation equipment</i>	1,076 ^d	2,801 ^d	N/A ^f
<i>ASU</i>	28,288 ^e	N/A ^f	N/A ^f
CO ₂ compressor	54,647 ^d	35,423 ^d	N/A ^f
ESP	1,259	1,259	1,259
FGD	0	7,500	7,500
SCR	2,750	2,750	2,750
Total	150,023	82,776	34,683
Overall Performance			
Steam turbine power, MW	717.1	440.9	533.2
Auxiliary power, MW	150.0	82.8	34.7
Heat input of coal, MW thermal (HHV)	1,319	1,318	1,318
Heat input of natural gas, MW thermal (HHV)	524	N/A ^f	N/A ^f
Net output power, MW	567.1	358.1	498.5
Net efficiency, % HHV	30.8	27.2	37.8

Table 10 (cont.)

^a Auxiliary load definitions:

ASU = Air Separation Unit

Ash handling = equipment required for conveying, preparing, storing, and disposing the flyash and bottom ash produced on a daily basis by the boiler

Coal handling = equipment required for unloading, conveying, preparing, and storing the coal delivered to the plant

Circulating water pumps = supply cooling water to condense the main turbine exhaust steam

Cooling tower fans = fans used to induce air flow within the cooling tower to cool warm water from the steam condenser

ESP = Electrostatic Precipitator

FGD = Flue Gas Desulfurization

Forced draft fans = fans that take air from the atmosphere and, first warming it in the air preheater, inject it via the air nozzles on the furnace wall

Induced draft fans = fans that assist the forced draft fans by drawing out flue gases from the furnace, maintaining a slightly negative pressure in the furnace to avoid backfiring through any opening

Induced draft fans (LSCC or MEA) = fans that restore pressure loss from the LSCC process (carbonator and calciner) or MEA process

Main feed pump = pumps feedwater from the deaerator storage tank to the economizer inlet on the boiler and to the LSCC heat exchangers. Includes smaller pump that pumps water from steam condenser to the deaerator. The deaerator storage tank assists in maintaining low levels of dissolved oxygen in the feedwater, while providing an enhanced ability to tolerate certain transients, such as trip of a condensate pump

Primary air fans = fans that take air from the atmosphere and provide the pressure to pass the air through the air pre-heater and coal pulverizers to the boiler burners

SCR = Selective Catalytic Reduction

Steam condensate pumps = pumps that pump condensate from the condenser hotwell through the feedwater heaters

Transformer loss = electricity loss due to the use of transformers to step the voltage up from the electricity generator as needed for transmission to its destination

Transportation equipment (LSCC or MEA) = pneumatic transportation of CaO using boiler flue gas and CaCO₃ using CO₂-rich gas stream in the LSCC case, and transportation of MEA using pumps in the MEA case

^b Energy use for particular unit operations in the LSCC and MEA processes were assumed to be the same as those in the base plant, unless indicated otherwise

^c For the MEA and Base cases, the portion of steam used to drive the turbine for water feeding is not counted in the gross output

^d Values were simulated in ChemCAD

^e Values were calculated (see Appendices B and C)

^f N/A = Not Applicable

^g HHV = Higher Heating Value

Process conditions, including consumables, sorbent mass flows, and operating temperatures and pressures, are shown in Table 7 for the LSCC and MEA process simulations. The energy usage for the ASU in the LSCC case was calculated based on a value of 160 kWh/short ton (176 kWh/metric ton) O₂ (see Appendix C4). Fans, pumps and the CO₂ product gas compressor operate at an assumed efficiency of 85%. The CO₂ product compressor consists of three stages, with an inter-stage cooling temperature of 40 °C.

While the major difference between the LSCC and MEA processes is the use of a solid sorbent instead of a liquid solvent to capture CO₂, there are several other significant differences that should be noted. The temperature in the carbonator for LSCC was maintained at 650 °C, compared to 40 - 57 °C for the absorber in the MEA process. This high temperature in the carbonator allowed the recovery of reaction heat to generate HP steam. Another significant difference between the two processes is that FGD for SO₂ removal is not necessary with the LSCC process. Finally, lime and limestone solids are transported in the LSCC process by using process gas streams (see Appendices C3 and C4), whereas the MEA solution is transported using liquid pumps.

Maintaining the calciner in the LSCC process at 950 °C required the combustion of natural gas and high-purity oxygen (direct heat exchange), whereas maintaining the stripper in the MEA process at 95 - 116 °C required the consumption of LP steam withdrawn from the power plant steam cycle (indirect heat exchange, where steam in the reboiler is not in direct contact with CO₂-laden MEA). Also, in the LSCC case a portion of heat is recovered from the CO₂-rich stream out of the calciner (950 °C), to produce additional HP steam for the plant. The limestone bleed (removal from the process) was assumed at zero for a “perfect” adsorption/desorption cycle (reflecting a best-case study for the LSCC process), whereas MEA solution bleed was 1 wt % of the regenerator MEA solution flow rate. Finally, the LSCC process in this study used the stoichiometric Ca/CO₂ molar ratio (1:1), which again presents a best-case study for LSCC, while the MEA process used the practical molar ratio (MEA/CO₂ = 5:1).

Regarding final stack emissions, the LSCC process occurs *before* the standard flue gas cleaning processes (SCR, ESP and FGD), while the MEA process occurs *after* these flue

gas cleaning processes. The SO₂ and HCl emissions were reduced to essentially zero – from the base plant simulation – with both the LSCC and MEA processes. While stack emissions of NO_x were also reduced to essentially zero with the MEA process, the LSCC process was assumed to not have affected these emissions. The stack CO₂ emissions with the LSCC and MEA processes are the same (90% removal of CO₂ from the inlet flue gas). The stack flow rates of other pollutants (PM and mercury), with both the LSCC and MEA processes, were assumed to not have changed from the base plant simulation.

The total auxiliary power use for the LSCC case is 150 MW_e, compared to 175 MW_e for MEA (which includes a 92 MW_e power loss due to steam extraction for MEA regeneration). The largest contributor to the power use in the LSCC case is compression of the CO₂ product gas from the LSCC process, which requires 55 MW_e. This is higher (+19 MW_e) compared to the MEA process due to a larger volumetric flow rate of CO₂ stream (from combustion of natural gas in calciner) and lower CO₂ pressure prior to compression in the LSCC process (15.8 psia vs. 25.0 psia, or 1.08 atm vs. 1.70 atm). The ASU in the LSCC case uses 28 MW_e of the available power to the power plant. The ASU produces oxygen, in order to combust natural gas for providing the heat in the calciner. Compared to the MEA case, the large amount of additional energy (+10 MW_e) consumed by the main feed pump in the LSCC case is a result of the extra water demand for heat recovery from the carbonator and calciner.

The net power output for the LSCC process scenario is 567 MW_e, compared to 358 MW_e for the MEA process scenario. A significant portion of the additional power output in the LSCC case is from heat recovered from the exothermic reaction in the carbonator (114 MW_e), as well as from the hot (650 °C) flue gas stream out of the carbonator (34 MW_e) and the high-temperature (950 °C) CO₂/H₂O stream out of the calciner (36 MW_e). The net efficiency for the LSCC process scenario is 30.8%, which is higher than the MEA process scenario (27.2%) but lower than the base scenario (37.8%).

As indicated in Section 1.4.2, select studies of limestone usage in the carbonation-calcination process for CO₂ capture at an air-blown PC power plant indicate the need for energy use details. The study by Ramkumar *et al*, using a hydrated lime process integrated with the rest of the plant, uses steam generated from the boiler flue gas as a

heat source at the calciner.¹⁸ However, the LSCC process here uses natural gas as a heat source for direct heat exchange at the calciner. The hydrated lime study resulted in an auxiliary power use of 104 MW_e (51 MW_e for CO₂ compression), and a net plant efficiency of 33%.¹⁸ The DOE/NETL target of a 30 - 50% reduction from the regeneration heat duty of the MEA process, for CO₂ removal processes utilizing solid sorbents for up to 90% CO₂ capture, results in an equivalent auxiliary energy use of 46 - 64 MW_e.²⁰ This auxiliary energy use does not include energy use for compression, pumps and fans, and is not comparable to the LSCC process here. The investigation of the limestone adsorption-desorption process by the European Commission targets a minimum CO₂ capture efficiency of only 60% from the boiler's flue gas for retrofitted power plants.²² Thus an energy use comparison with the European Commission process is not applicable for the LSCC process here which captures 90% CO₂ from the boiler flue gas. The simulated process by Grasa *et al*, which included an analysis of heat recovery from the hot (1000 °C) CO₂ stream out of the desorption reactor, was based on a 1000 MW_{th} plant and removed only 33% of the CO₂ in the boiler flue gas.¹⁹ Regarding energy use, this study also is not comparable to the LSCC process here.

4 SUMMARY OF CONCLUSIONS AND RECOMMENDATIONS AND FUTURE WORK

4.1 IECM Evaluations

Recent regulations from the U.S. Environmental Protection Agency limit emissions of sulfur dioxide (SO₂), nitrogen oxides (NO_x), particulate matter (PM) and mercury to the atmosphere. As a result of these regulations, this study used the Integrated Environmental Control Model (IECM) to investigate the economic competitiveness of burning select blends of Illinois high-sulfur bituminous (or, “IL”) and western low-sulfur Powder River Basin (PRB) coals at pulverized coal (PC) power plants while meeting the air emission limitations in these regulations. IECM allows for specifying fuel cost, fuel composition, and environmental compliance limits for SO₂, NO_x, PM and mercury. Assumptions for particular IECM performance requirements were developed and used as inputs in IECM. These inputs were used to evaluate scenarios with different coal feed types and environmental controls. Model parameters tested were: coal type and cost; control technology for SO₂, NO_x, PM and mercury; and boiler size. The breakeven costs of electricity with respect to use of different fuels, as well as the capital and operation and maintenance costs, were estimated for the different scenarios.

For a 650 MW_e (gross) PC power plant, 100% Illinois #6 high-sulfur bituminous coal costing more than \$33.0/short ton (\$36.4/metric ton) could not compete with 100% western PRB low-sulfur sub-bituminous coal at the 2007 PRB coal cost of \$27.5/short ton (\$30.3/metric ton). However, it was determined that a 70/30 PRB/IL coal blend (weighted-averages of the two individual coals according to their ratio in the blend) at the 2007 market costs of Illinois coal (\$35.0/short ton, or \$38.6/metric ton), as well as Illinois coal at a new Illinois mine-mouth facility (\$17.6/short ton, or \$19.4/metric ton), can be competitive with 100% PRB at \$27.5/short ton (\$30.3/metric ton). The 100% Illinois coal case becomes less competitive with a decrease in plant size, and for a 175 MW_e (gross) PC plant even low-cost Illinois coal (\$30.0/short ton, or \$33.1/metric ton) does not compete with 100% PRB coal or the 70/30 PRB/IL coal blend. *The most cost-effective case scenario for PC plants, at the 2007 market coal costs (\$27.5/short ton (\$30.3/metric ton) for PRB coal, \$35.0/short ton (\$38.6/metric ton) for Illinois coal, and \$17.6/short*

ton (\$19.4/metric ton) for Illinois mine-mouth coal), is the Illinois mine-mouth for a 650 MW_e (gross) plant (\$67.0/MWh) and the 70/30 PRB/IL coal blend for a 175 MW_e (gross) plant (\$95.0/MWh). The Illinois mine-mouth coal is most cost-effective for the 650 MW_e case due primarily to its lower coal cost compared to the other coal types. The lower coal cost helps compensate for the higher SO₂ control cost for the Illinois mine-mouth coal compared to the other coal types. The 70/30 PRB/IL coal blend replaces the Illinois mine-mouth as most cost-effective when the plant size is reduced to 175 MW_e – even though the Illinois mine-mouth coal still has a lower coal cost compared to the other coal types – for the following reason: the SO₂ and PM control costs for the Illinois mine-mouth case have a higher contribution to plant COE at the 175 MW_e plant than at the 650 MW_e plant, whereas these control costs for the 70/30 PRB/IL coal blend have essentially the same contribution to plant COE at both plant sizes.

Since power plants are one of the biggest emission sources of carbon dioxide (CO₂) – a main greenhouse gas – and thus a target for potential CO₂ regulations, 90% CO₂ capture from the flue gas using a monoethanolamine (MEA) absorption process was also evaluated within the IECM study. With the addition of CO₂ capture using the MEA process, the breakeven delivered Illinois coal costs increase for 100% Illinois coal and the 70/30 PRB/IL coal blend, making these coals more competitive when compared to the 100% PRB coal case. However, the breakeven delivered Illinois coal cost decreases with the addition of CO₂ capture for the Illinois mine-mouth coal case, making the Illinois mine-mouth coal less competitive when compared to the 100% PRB coal case. For a 650 MW_e (gross) PC plant with CO₂ capture, 100% Illinois coal costing less than \$36.0/short ton (\$39.7/metric ton) can be cost-effective relative to the 100% PRB coal case at \$27.5/short ton (\$30.3/metric ton). Also with addition of CO₂ capture, the 100% Illinois coal case (at the 2007 cost of \$35.0/short ton, or \$38.6/metric ton) can compete with the following cases: 100% PRB coal at PRB coal costs of \$27.5 and \$32.0/short ton (\$30.3 and \$35.3/metric ton); and the 70/30 PRB/IL coal blend at a PRB coal cost of \$32.0/short ton (\$35.3/metric ton) for the 650 MW_e (gross) plant. As with the PC cases without CO₂ capture, the 100% Illinois coal case with the addition of CO₂ capture becomes less competitive with a decrease in plant size. The 100% Illinois coal case (at the 2007 market cost of \$35.0/short ton, or \$38.6/metric ton) cannot compete with either 100% PRB coal

or the 70/30 PRB/IL coal blend at PRB coal costs of \$20.0, \$27.5 and \$32.0/short ton (\$22.0, \$30.3 and \$35.3/metric ton) for the 175 MW_e (gross) case. Emissions of SO₂, NO_x and PM (short tons/hour and metric tons/hour) decreased with the addition of CO₂ capture to a PC plant, while mercury emissions (short tons/hour and metric tons/hour) remained the same. *The most cost-effective case scenario for PC plants without CO₂ capture, at the 2007 market coal costs (\$27.5/short ton (\$30.3/metric ton) for PRB coal, \$35.0/short ton (\$38.6/metric ton) for Illinois coal, and \$17.6/short ton (\$19.4/metric ton) for Illinois mine-mouth coal) is the Illinois mine-mouth for a 650 MW_e (gross) plant (\$67.0/MWh) and the 70/30 PRB/IL coal blend for a 175 MW_e (gross) plant (\$95.0/MWh). The most cost-effective case scenario for PC plants seeking to add CO₂ capture, at the 2007 market coal costs, is the 70/30 PRB/IL coal blend for both a 650 MW_e (gross) plant (\$133.1/MWh) and a 175 MW_e (gross) plant (\$177.0/MWh). The CO₂ control cost for the 70/30 PRB/IL coal blend is higher than the 100% Illinois and Illinois mine-mouth coals, due to the higher CO₂ emissions inherent to burning PRB coal. However, the higher SO₂ control costs for the 100% Illinois and Illinois mine-mouth coals, compared to the 70/30 PRB/IL coal blend, counter the lower CO₂ control costs for these coals sufficiently to make the 70/30 PRB/IL coal blend most cost-effective for the 650 MW_e and 175 MW_e cases.*

For future work, it is recommended that a 70/30 PRB/IL coal blend be fired in a field study to confirm the conclusions of this study. The field study should demonstrate that select blends of coal can be fired in a boiler while using dry flue gas desulphurization, selective catalytic reduction, and fabric filtration with 90% removal of mercury emissions without problems for the operator of the boiler.

4.2 LSCC Process Simulation

Most existing CO₂ capture systems utilize absorption-based technology, though it is an energy intensive process. The solid adsorption method has potential to be competitive with the MEA absorption process, regarding energy conversion efficiency for a PC power plant. Therefore, a second part of this study investigated a process using limestone to remove 90% of the CO₂ emissions at a PC power plant. The configuration and operating conditions of a 533 MW_e (gross) U.S. Department of Energy (DOE) reference PC power

plant (referred to as the “base” plant) were modified to accommodate inclusion of the limestone carbonation-calcination (LSCC) process for capturing CO₂ within the boiler flue gas. A 533 MW_e (gross) power plant equipped with the MEA process for CO₂ capture, also based on the U.S. DOE reference PC power plant, was included for comparison. Energy usage for the different unit operations at the plants with LSCC and MEA were assumed to be the same as the base plant, estimated from the base plant, or found by direct simulation. Simulation of the LSCC process was performed using ChemCAD (version 6.1.0), as a “proof-of-concept” study with the goal of estimating the best-case energy use of the process at a PC power plant. Process inputs – for the combustion, steam generation, carbonator, calciner, and CO₂ product sections of the plant – were used in the ChemCAD program to obtain results of auxiliary power use and overall process performance.

The total auxiliary power use for the LSCC case is 150 MW_e, compared to 175 MW_e for MEA (which includes a 92 MW_e power loss due to steam extraction for MEA regeneration). The power use in the LSCC case is attributed primarily to the CO₂ compressor (55 MW_e), followed by the air separation unit (ASU) at the calciner (28 MW_e) and the main feed pump (21 MW_e). The power use for CO₂ compression in the LSCC process is higher than the MEA process, due to a larger volumetric flow rate of CO₂ stream (from combustion of natural gas in the calciner) and lower CO₂ pressure prior to compression in the LSCC process. The ASU produces high-purity oxygen for combustion of natural gas, which provides a heat source for the calciner. Additional energy is consumed by the main feed pump in the LSCC process compared to the MEA process, due to the extra water demand for heat recovery from the carbonator and calciner. Also, flue gas desulfurization for SO₂ removal is not necessary with the LSCC process, which further reduces power use. High quality carbonation reaction heat, heat from the CO₂-lean flue gas at the heat exchanger after the carbonator, and heat from the CO₂/water vapor gas stream at the heat exchanger after the calciner are recovered for producing steam used for additional electricity generation in the LSCC process. The recovered heat in the LSCC process is associated with high temperatures in the carbonator (650 °C) and calciner (950 °C), whereas the MEA regenerator consumes low-pressure steam (60 psia, or 4.1 atm, and 246 °C) from the power plant steam cycle. The

additional electricity leads to a higher net generation efficiency for the LSCC process (30.8%) compared to the MEA process (27.2%). *Therefore, when considering energy requirement for CO₂ capture at a PC plant, the LSCC process is more favorable than the MEA process. It should be noted that this is for the best-case scenario of LSCC based on ideal operating conditions.*

For future work, the following factors for the LSCC process should be evaluated to better assess the overall competitiveness with the MEA process with respect to energy use, process conditions, and emissions: Ca/CO₂ molar ratios of 3:1 and 5:1; use of voidage values from 0.95 to 0.98 for fluidization of particles in the carbonator and calciner for pressure differential calculations; and use of coal instead of natural gas for heat supply at the calciner.

REFERENCES

1. International Energy Outlook 2009, 2009. U.S. Energy Information Administration Analysis and Projections Web site. [http://www.eia.doe.gov/oiaf/archive/ieo09/pdf/0484\(2009\).pdf](http://www.eia.doe.gov/oiaf/archive/ieo09/pdf/0484(2009).pdf) (accessed Sep 15, 2009).
2. Annual Energy Outlook 2008, 2008. U.S. Energy Information Administration Analysis and Projections Web site. [http://www.eia.doe.gov/oiaf/aeo/pdf/0383\(2008\).pdf](http://www.eia.doe.gov/oiaf/aeo/pdf/0383(2008).pdf) (accessed Jul 8, 2009).
3. U.S. Energy Information Administration International Energy Statistics: Consumption. <http://tonto.eia.doe.gov/cfapps/ipdbproject/IEDIndex3.cfm?tid=1&pid=1&aid=2> (accessed Oct 8, 2009).
4. U.S. Energy Information Administration Sources and Uses: Coal. <http://www.eia.doe.gov/fuelcoal.html> (accessed Aug 25, 2007).
5. Illinois Administrative Code, Title 35, Subtitle B, Chapter 1, Subchapter c, Part 225 Control of Emissions from Large Combustion Sources, 2005. Illinois Pollution Control Board Web site. <http://www.ipcb.state.il.us/documents/dsweb/Get/Document-55740/> (accessed June 4, 2007).
6. Title 40, Part 60, Subpart D Standards of Performance for Fossil-Fueled Steam Generators for which Construction is Commenced after August 17, 1971, 2007. National Archives and Records Administration Electronic Code of Federal Regulations Web site. <http://ecfr.gpoaccess.gov/cgi/t/text/textidx?c=ecfr&sid=22104e17f3384a3e823e978150aed838&rgn=div6&view=text&node=40:6.0.1.1.1.9&idno=40> (accessed Aug 17, 2007).
7. National Oceanic and Atmospheric Administration. Earth System Research Laboratory: Global Monitoring Division. <http://www.esrl.noaa.gov/gmd> (accessed August 15, 2011).
8. International Energy Outlook 2008, 2008. U.S. Energy Information Administration Analysis and Projections Web site. [http://www.eia.doe.gov/oiaf/archive/ieo08/pdf/0484\(2008\).pdf](http://www.eia.doe.gov/oiaf/archive/ieo08/pdf/0484(2008).pdf) (accessed January 15, 2009).
9. Quadrelli, R.; Peterson, S. The Energy-Climate Challenge: Recent Trends in CO₂ Emissions from Fuel Combustion. *Energy Policy* **2007**, *35*, 5940.

10. Inventory of U.S. Greenhouse Gas Emission Sources and Sinks: 1990-2009, 2011.
U.S. Environmental Protection Agency Climate Change Web site. <http://epa.gov/climatechange/emissions/usinventoryreport.html> (accessed Feb 25, 2012).
11. Greenhouse Gas Legislation: Summary and Analysis of H.R. 2454 as Passed by the House of Representatives, 2009. Congressional Research Service Web site.
<http://opencrs.com/document/R40643/> (accessed Oct 15, 2010).
12. Strategic Plan for the IEA Demand-Side Management Programme 2008-2012.
International Energy Administration Demand Side Management Programme
Web site. <http://www.ieadsm.org/Files/Exco%20File%20Library/Participation/Final%20strategy%202008-2012.pdf> (accessed Oct 30, 2010).
13. CO₂ Capture and Storage: A Key Carbon Abatement Option, 2008. International Energy Agency Web site. http://www.iea.org/textbase/nppdf/free/2008/CCS_2008.pdf (accessed Oct 29, 2010).
14. Chen, S.; Lu, Y.; Rostam-Abadi, M. *Carbon Dioxide Capture and Transportation Options in the Illinois Basin*. Topical Report to U.S. Department of Energy on U.S. DOE Cooperative Agreement DE-FC26-03NT41994; Illinois State Geological Survey: Champaign, Illinois, 2005.
15. Abanades, J. C.; Rubin, E. S.; Anthony, E. J. Sorbent Cost and Performance in CO₂ Capture Systems. *Ind. Eng. Chem. Res.* **2004**, *43*, 3462-3466.
16. Davidson, R. *Post-Combustion Carbon Capture – Solid Sorbents and Membranes*. Report; International Energy Agency Clean Coal Centre: London, 2009.
17. Integrated Environmental Control Model: A Tool for Calculating the Performance, Emissions, and Cost of a Fossil-Fueled Power Plant Version 6.2.4, 2010.
Carnegie Mellon University Engineering and Public Policy Web site. <http://www.cmu.edu/epp/iecm/> (accessed Oct 20, 2010).
18. Ramkumar, S.; Wang, W.; Li, S.; Gumuluru, S.; Sun, Z.; Phalak, N.; Wong, D.; Iyer, M.; Statnick, R. M.; Fan, L.-S.; Sakadjian, B. Carbonation-Calcination Reaction (CCR) Process for High Temperature CO₂ and Sulfur Removal. Presented at 1st Meeting of the High Temperature Solid Looping Cycles Network. [Online] September 15-17, 2009. Columbus, Ohio. IEAGHG High Temperature Looping Cycles Network. <http://www.ieaghg.org/docs/loopingpdf/16>

- %20septiembre/C13.pdf (accessed Jun 15, 2010).
19. In-Duct Sorbent Injection for CO₂ Capture, 2004. U.S. Department of Energy National Energy Technology Laboratory Web site. <http://www.netl.doe.gov/publications/proceedings/04/carbon-seq/155.pdf> (accessed Jul 30, 2010).
 20. Reactor Design for CO₂ Capture Using Sorbents, 2010. U.S. Department of Energy National Energy Technology Laboratory Web site. <http://www.netl.doe.gov/publications/factsheets/rd/R%26D123.pdf> (accessed Nov 1, 2010).
 21. Evaluation of Solid Sorbents as a Retrofit Technology for CO₂ Capture from Coal-Fired Power Plants. Presented at the Annual NETL CO₂ Capture Technology for Existing Plants R&D Meeting. [Online] March 24, 2009. Pittsburgh, Pennsylvania. U.S. Department of Energy National Energy Technology Laboratory Web site. <http://www.netl.doe.gov/technologies/coalpower/ewr/co2/post-combustion/solid-sorbent.html> (accessed Jun 20, 2010).
 22. CO₂ Capture and Storage Projects, 2009. European Commission Research and Innovation Web site. http://ec.europa.eu/research/energy/pdf/synopses_co2_en.pdf (accessed Jul 31, 2010).
 23. Chen, S.; Loftus, J. P.; Lu, Y.; Statnick, R. M. *Economic Evaluation of Illinois Coal and Western PRB Coal under Different Pollution Control Scenarios*. Final Technical Report to Illinois Clean Coal Institute on ICCI Cooperative Agreement 05-1/US-2; Illinois State Geological Survey: Champaign, Illinois, 2007.
 24. Rubin, E. S. Implications of Future Environmental Regulation of Coal-Based Electric Power. *Annu. Rev. Energy* **1989**, *14*, 19-45.
 25. Penner, S.S.; Alpert, S.B.; Beár, J.N.; Denn, M.; Haag, W.; Magee, R.; Reichl, E.; Rubin, E. S.; Solomon, P.R.; Wender, I.; Woodcock, K. Coal Gasification: Direct Applications and Syntheses of Chemicals and Fuels. *Pergamon Journals* **1987**.
 26. Rubin, E. S.; Cushey, M. A.; Marnicio, R. J.; Bloyd, C. N.; Skea, J. F. Controlling Acid Deposition: The Role of FGD. *Environ. Sci. Technol.* **1986**, *20* (10), 960-969.
 27. Rubin, E. S.; Kalagnanam, J. R.; Frey, H. C.; Berkenpas, M. B. Integrated Environmental Control Modeling of Coal-Fired Power Systems. *J. Air Waste Manage. Assoc.* **1997**, *47*, 1180-1186.

28. Frey, H. C.; Rubin, E. S. Evaluation of Advanced Coal Gasification Combined-Cycle Systems under Uncertainty. *Ind. Eng. Chem. Res.* **1992**, *31*, 1299-1307.
29. Chow, W. C.; Miller, M. J.; Fortune, J.; Behrens, G.; Rubin, E. S. Managing Air Toxics Under the New Clean Air Act Amendments. *Power Engineering* **1991**, Jan, 30-34.
30. Rubin, E. S.; Chen, C.; Rao, A. B. Cost and Performance of Fossil Fuel Power Plants with CO₂ Capture and Storage. *Energy Policy* **2007**, *35* (9), 4444-4454.
31. Rao, A. B.; Rubin, E. S. A Technical, Economic, and Environmental Assessment of Amine-Based CO₂ Capture Technology for Power Plant Greenhouse Gas Control. *Environ. Sci. Technol.* **2002**, *36* (20), 4467-4475.
32. *Evaluation of Power Generators in the Illinois Basin under CO₂ Control Scenarios*. Final Technical Report to Illinois Clean Coal Institute on ICCI Cooperative Agreement 05-1/US-2 Subtask 16.2; Illinois State Geological Survey: Champaign, Illinois, 2007.
33. Integrated Environmental Control Model: A tool for calculating the performance, emissions, and cost of a fossil-fueled power plant, Version 5.21, 2007. Carnegie Mellon University Engineering and Public Policy Web site. www.iecm-online.com (accessed April 15, 2011).
34. Withum, J. A. *Evaluation of Mercury Emissions from Coal-Fired Facilities with SCR and FGD Systems*. Final Report to U.S. Department of Energy on U.S. DOE NETL Cooperative Agreement DE-FC26-02NT41589; CONSOL Energy Inc.: South Park, Pennsylvania, 2006.
35. Sjoström, S. Evaluation of Sorbent Injection for Mercury Control. Presented at U.S. Department of Energy National Energy Technology Laboratory Meeting. [Online] July, 2009. U.S. DOE National Energy Technology Laboratory Web site. <http://www.netl.doe.gov/publications/proceedings/05/Mercury/pdf/Sjoström-071205-am.pdf> (accessed Aug 17, 2007).
36. Lee, J. Y.; Keener, T. C.; Yang, Y. J. Potential Flue Gas Impurities in Carbon Dioxide Streams Separated from Coal-Fired Power Plants. *J. Air Waste Manage. Assoc.* **2009**, *59*, 725-732.

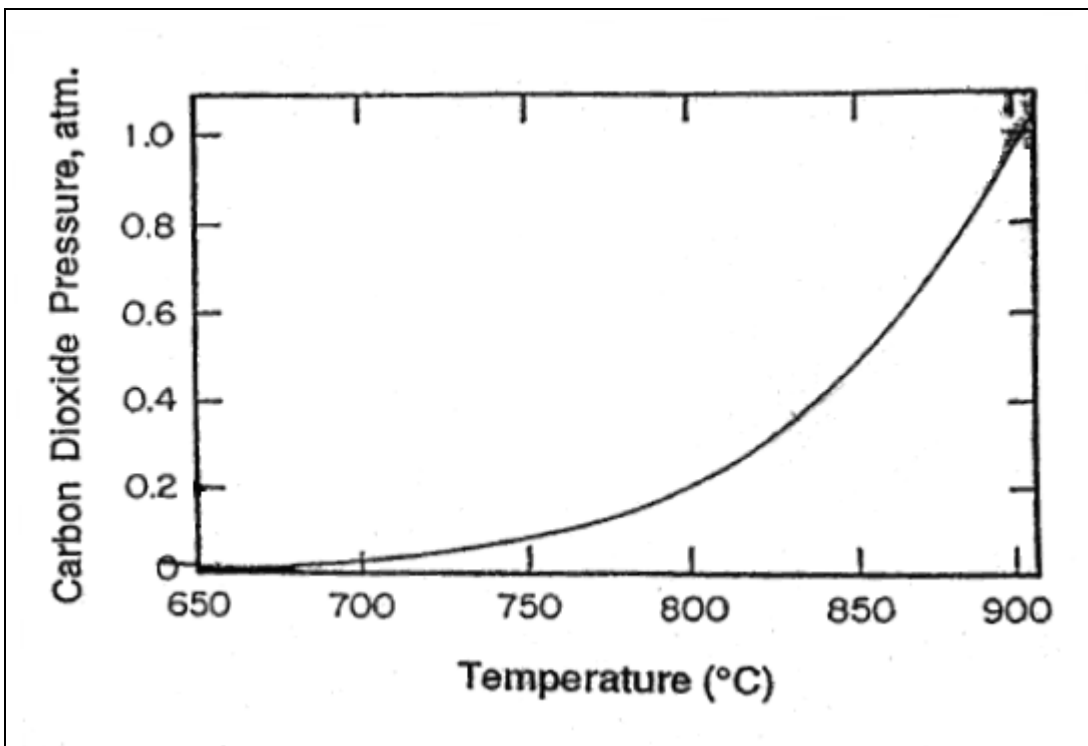
37. *Clean Coal Reference Plants: Pulverized Coal Boiler With Flue Gas Desulfurization*. Topical Report to U.S. Department of Energy on U.S. DOE Cooperative Agreement DE-AM21-94MC31166; Gilbert/Commonwealth, Inc.: Reading, Pennsylvania, 1995.
38. Squires, A. M. Cyclic Use of Calcined Dolomite to Desulfurize Fuels Undergoing Gasification. *Adv. Chem. Ser.* **1967**, 69, 205-229.
39. Curran, G. P.; Fink, C. E.; Gorin, E. Carbon Dioxide-Acceptor Gasification Process. Studies of Acceptor Properties. *Adv. Chem. Ser.* **1967**, 69, 141-165.
40. Abanades, J. C.; Anthony, E. J.; Wang, J.; Oakley, J. E. Fluidized Bed Combustion Systems Integrating CO₂ Capture with CaO. *Environ. Sci. Technol.* **2005**, 39, 2861-2866.
41. Abanades, J. C.; Anthony, E. J.; Lu, D. Y.; Salvador, C.; Alvarez, D. Capture of CO₂ from Combustion Gases in a Fluidized Bed of CaO. *AIChE J.* **2004**, 50 (7), 1614-1622.
42. Shimuzu, T.; Hirama, T.; Hosoda, H.; Kitano, K.; Inagaki, M.; Tejima, K. A Twin Fluid-Bed Reactor for Removal of CO₂ from Combustion Processes. *Trans. IChemE.* **1999**, 77, 62-68.
43. Statnik, R. M. Clear Skies Consulting, LLC, Charlotte, North Carolina. Personal communication, 2007.
44. Levasseur, A. A. Alstom's Oxy-Firing Technology Development and Demonstration - Near Term CO₂ Solutions. Presented at the 34th International Technical Conference on Clean Coal & Fuel Systems. [Online] June, 2009. Clearwater, Florida. U.S. Department of Energy National Energy Technology Laboratory Web site. <http://www.netl.doe.gov/technologies/coalpower/ewr/co2/pubs/5290%20Alstom%20oxy-combustion%20paper%20Clearwater%20jun09.pdf> (accessed Mar 17, 2010).
45. Alie, C.; Backham, L.; Croiset, E.; Douglas, P. L. Simulation of CO₂ Capture using MEA Scrubbing: A Flowsheet Decomposition Method. *Energy Convers. Manage.* **2005**, 46, 475-487.
46. Freguia, S.; Rochelle, G. Modeling of CO₂ Capture by Aqueous Monoethanolamine. *AIChE J.* **2003**, 49 (7), 1676-1686.

47. White, C.W. *Aspen Plus Simulation of CO₂ Recovery Process*. Technical Report to U.S. Department of Energy on U.S. DOE Cooperative Agreement DOE/NETL-2002/1182: Pittsburgh, PA, September 2003.
48. Grasa, G. S.; Abanades, J. C. CO₂ Capture Capacity of CaO in Long Series of Carbonation/Calcination Cycles. *Ind. Eng. Chem. Res.* **2006**, *45*, 8846-8851.
49. Abanades, J. C.; Alvarez, D. Conversion Limits in the Reaction of CO₂ with Lime. *Energy Fuels* **2003**, *17*, 308-315.
50. Farzan, H.; Figueroa, J. Developing Oxy-combustion for Retrofitting Coal-fired Boilers. Presented at 7th Annual Conference on Carbon Capture and Sequestration. [Online] May 2008. Pittsburgh, PA. <http://www.netl.doe.gov/technologies/coalpower/ewr/co2/pubs/42747%20B%26W%20oxy-combustion%207th%20CCS%20conference%202008.pdf> (accessed Feb 20, 2011).
51. Lu, Y. Illinois State Geological Survey, Champaign, IL. Personal communication, 2007.
52. Abbasian, J.; Hassanzadeh, A. Development of Highly Durable and Reactive Regenerable Magnesium-Based Sorbents for CO₂ Separation in Coal Gasification Processes. Presented at Annual UCR Contractors' Review Meeting. [Online] June 10, 2004. Pittsburgh, PA. http://www.netl.doe.gov/publications/proceedings/04/ucrhcuc/presentations/Abbasian_P.pdf (accessed Mar 3, 2012).
53. Liang, Y. Carbon Dioxide Capture from Flue Gas Using Regenerable Sodium-Based Sorbents. M.S. Thesis, Louisiana State University, August 2003.
54. Lu, Y. Illinois State Geological Survey, Champaign, IL. Personal communication, 2011.
55. Fluidized Bed Boiler Designs, 2011. Greenfield Research Center Products Web site. <http://www.greenfieldresearch.ca/bedboiler.html> (accessed Jun 15, 2011).
56. Rhodes, M. Pneumatic Transport of Powders, 2001. Educational Resources for Particle Technologies Web site. <http://www.erpt.org/retiredsite/014Q/rhoe-00.htm> (accessed Mar 3, 2012).

APPENDICES

Appendix A General Items for LSCC Process

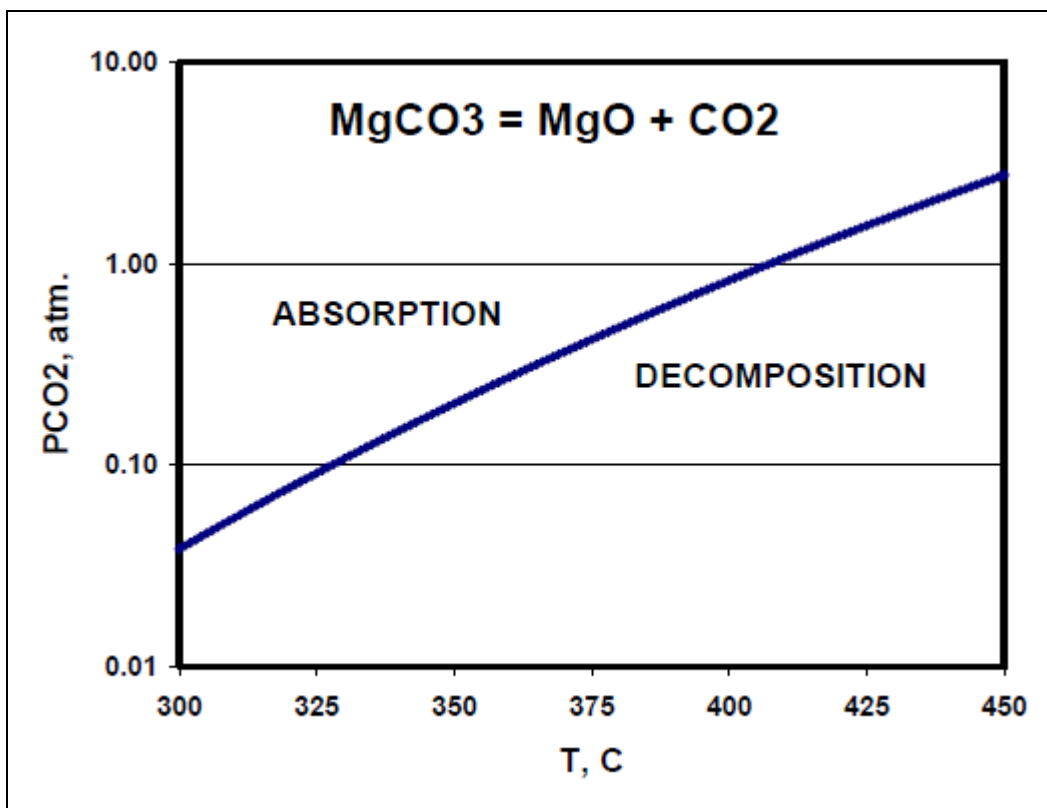
Appendix A1 Equilibrium Curve for CaO with CO₂



Note: Figure created in ChemCAD⁵¹

Figure 19 Equilibrium curve for CaO with CO₂

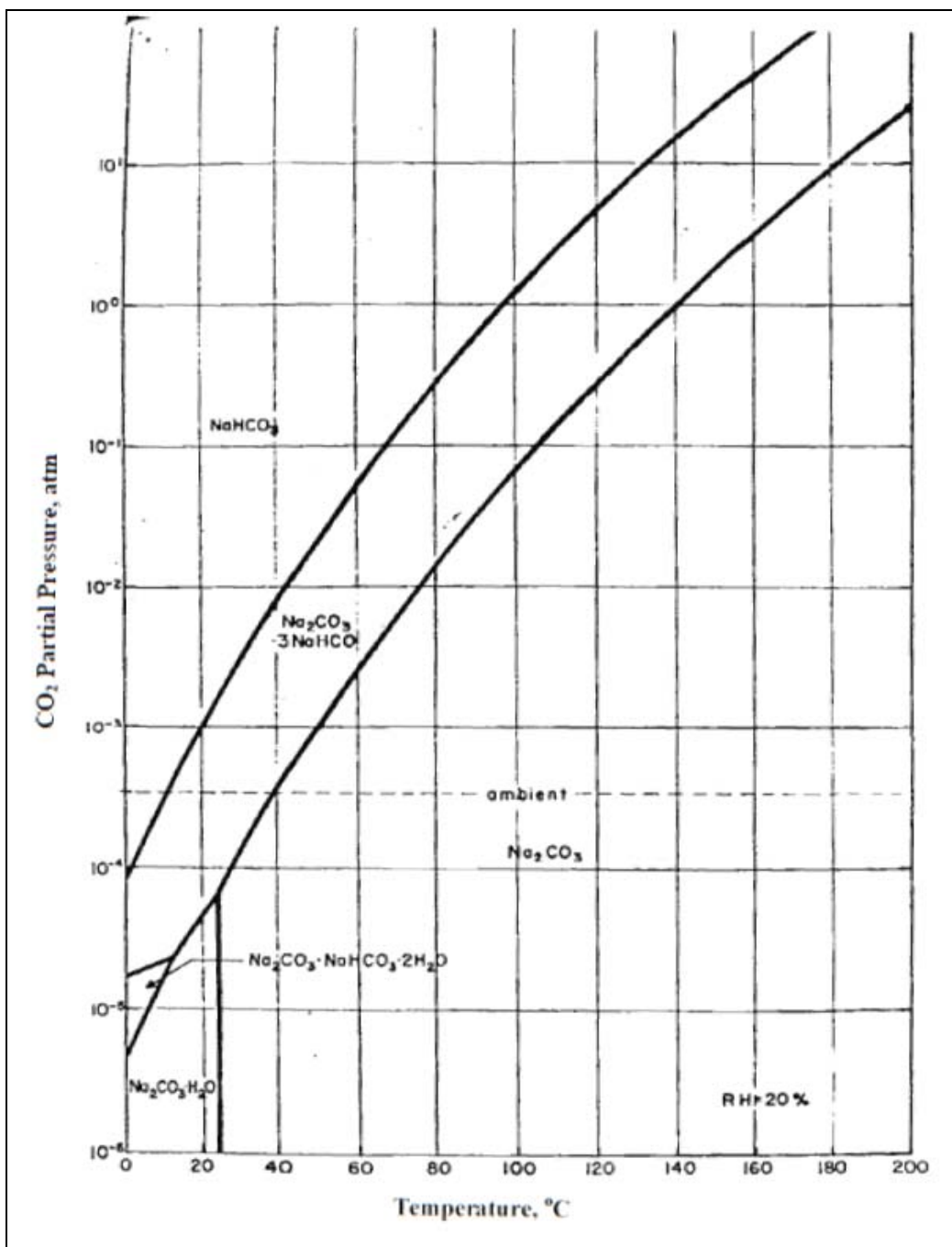
Appendix A2 Equilibrium Curve for MgO with CO₂



Note: Figure accessed from Illinois Institute of Technology⁵²

Figure 20 Equilibrium curve for MgO with CO₂

Appendix A3 Equilibrium Curve for Na_2CO_3 with CO_2



Note: Figure accessed from Y. Liang, Louisiana State University⁵³

Figure 21 Equilibrium curve for Na_2CO_3 with CO_2

Appendix A4 Steam Utilization Curve

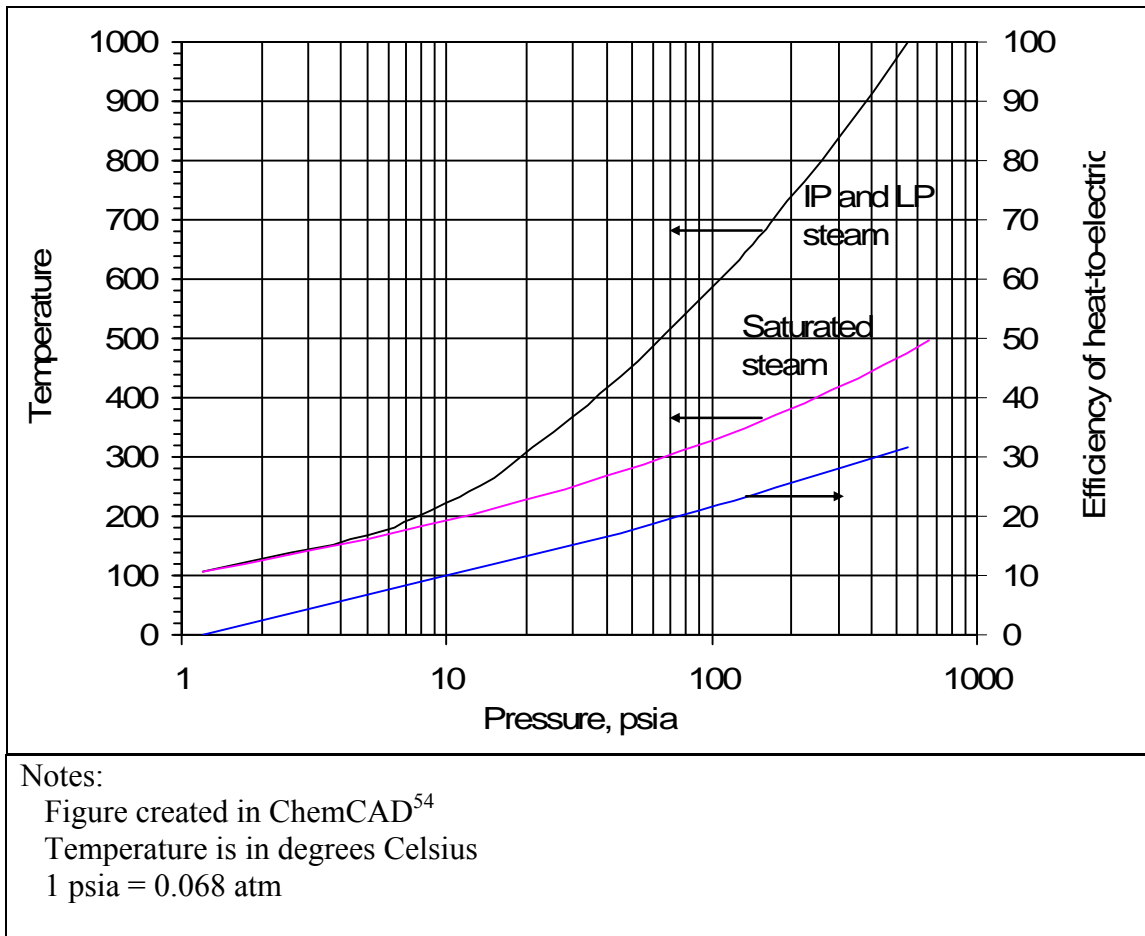


Figure 22 Steam utilization curve

Appendix B Assumptions for ChemCAD Simulations of Base Plant and Base Plant with MEA Process

Appendix B1 Operating Conditions of Base Plant

Table 11 Operating conditions of base plant

Parameter	PC plant with no CO ₂ control	Data source
Combustion		
Air/coal equivalent ratio	1.15	Reference ³⁷
Air flow rate, lb/hr (kg/hr)	3,991,198 (1,810,407)	Reference ¹⁴
Coal feed rate, lb/hr (kg/hr)	360,611 (163,573)	Reference ¹⁴
Steam generation		
Hot reheat steam, lb/hr (kg/hr)	3,022,125 (1,370,836)	Reference ¹⁴
Superheat steam, lb/hr (kg/hr)	3,422,824 (1,552,593)	
Steam condensate, lb/hr (kg/hr)	2,802,051 (1,271,010)	
Main feed-water, lb/hr (kg/hr)	3,321,228 (1,506,509)	
Heat duty of cooling tower, mmBtu/hr (MJ/hr)	2,178 (2,297,790)	
Flue gas (post-FGD)		
Flue gas flow rate, lb/hr (kg/hr)	4,472,755 (2,028,842)	Reference ¹⁴
Flue gas temperature, °C	54	Assumed ¹⁴
Composition, mol%		
N ₂	70.73	Reference ¹⁴
O ₂	2.47	
CO ₂	13.75	
H ₂ O	13.04	
SO ₂	0.0080	
NO _x	0.0032	
Fly ash flow rate, lb/hr (kg/hr)	257 (117)	Simulated ¹⁴

Appendix B2 Operating Conditions for MEA Process

Table 12 Simulation Parameters for MEA Process

Parameter	PC plant with MEA	Data source
Combustion		
Air/coal equivalent ratio	1.15	Reference ³⁷
Air flow rate, lb/hr (kg/hr)	3,991,198 (1,810,407)	Reference ¹⁴
Coal feed rate, lb/hr (kg/hr)	360,611 (163,573)	Reference ¹⁴
Steam generation		
Hot reheat steam, lb/hr (kg/hr)	3,022,125 (1,370,836)	Reference ¹⁴
Superheat steam, lb/hr (kg/hr)	3,422,824 (1,552,593)	
Steam condensate, lb/hr (kg/hr)	2,802,051 (1,271,010)	
Main feed-water, lb/hr (kg/hr)	3,321,228 (1,506,509)	
Heat duty of cooling tower, mmBtu/hr (MJ/hr)	1,104 (1,164,720)	
Absorber		
Inlet flue gas volume, lb/hr (kg/hr)	4,472,743 (2,028,836)	Reference ¹⁴
Gas inlet temperature, °C	54	Reference ^{45,46,47}
SO ₂ inlet, lb/hr (kg/hr)	16,458 (7,465)	Simulated
Active MEA solvent concentration, wt %	30	Reference ^{45,46,47}
Active MEA solvent/CO ₂ molar ratio	5:1	Reference ^{45,46,47}
Operating temperature, °C	40 - 57	Simulated ¹⁴
Pressure drop, psi (atm)	3.0 (0.20)	Assumed ¹⁴
CO ₂ capture efficiency, wt %	90	Assumed ¹⁴
CO ₂ captured, lb/hr (kg/hr)	834,748 (378,642)	Simulated ¹⁴
Stripper		
MEA solvent make-up, lb/hr (kg/hr)	2,794 (1,267)	Reference ¹⁴
LP steam consumption, lb/hr (kg/hr)	1,380,364 (626,133)	Reference ¹⁴
Stripping temperature, °C	95 - 116	Simulated ¹⁴
Pressure drop, psi (atm)	3.0 (0.20)	Assumed ^{45,46,47}
CO ₂ product		
Flow rate, lb/hr (kg/hr)	834,748 (378,642)	Simulated ¹⁴
Temperature, °C	20	Reference ^{45,46,47}
Pressure, psia (atm)	1,205 (82)	Reference ¹⁴

Appendix B3 Energy Use for Circulating Water Pumps and Cooling Tower Fans

1) Circulating Water Pump Energy Use

The energy use was calculated by linearly scaling up the base plant circulating water pumps energy use, based on the steam condensate flow rate from the base case simulation.

2) Cooling Tower Fans Energy Use

The energy use was calculated by linearly scaling up the base plant cooling tower fans energy use, based on cooling tower heat duty from the base case simulation.

Appendix C Assumptions for ChemCAD Simulation of LSCC Process

Appendix C1 Boiler

Table 13 Properties of Illinois coal¹⁴

Composition	As received (wt %)	Dry (wt %)	Dry, ash free (wt %)	Dry, ash free (mol/100g)	Number of atoms of each element in formula ^a	Combustion product	Product (mol)	Heat of formation (kJ/mol)	kJ	
Moisture	6.08	-	-	-						
Carbon, C	70.28	74.83	82.67	6.88	164.39	CO ₂	164.39	-393.52	-64,692	
Hydrogen, H	4.77	5.08	5.61	5.57	132.98	H ₂ O	66.41	-285.83	-18,982	
Nitrogen, N	1.44	1.53	1.69	0.12	2.88	NO	2.88	90.25	260	
Chlorine, Cl	0.21	0.22	0.24	0.01	0.16	-	-	-	0	
Sulfur, S	2.28	2.43	2.68	0.08	2.00	SO ₂	2.00	-296.83	-594	
Oxygen, O	6.25	6.65	7.35	0.46	10.97	-O ₂	5.48	0	0	
Ash	8.9	9.48	-	-	-	-	-	-	-	
					2,388.5 (molecular weight, g/mol)				-2,557	Coal heat of formation, kJ/mol ^c
HHV(Btu/lb) ^b	12,475	13,283	14,674		C ₁₆₄ H ₁₃₃ O ₁₁ N ₃ S ₂					
HHV(kJ/mol) ^b	69,244	73,729	81,450							
^a Calculated based on the simplest whole-number ratio of atoms using the dry, ash free values										
^b HHV = Higher Heating Value										

Table 13 (cont.)

^c Coal heat of formation was calculated as follows:

$$\begin{aligned}
 \text{Coal heat of formation} &= HHV_{\text{coal}} - \sum_j v_j + \sum_i v_i \\
 &= 81,450 \text{ kJ} - 84,267 \text{ kJ} + 260 \text{ kJ} \\
 &= -2,557 \text{ kJ per mol of coal}
 \end{aligned}$$

where v_j represents the heat of formation in kJ for each product j , and v_i represents the heat of formation in kJ for each reagent i with $i \neq \text{coal}$

Appendix C2 Operating Conditions for LSCC Process

Table 14 Simulation parameters for LSCC process

Parameter	PC plant with LSCC	Data source
Combustion		
Air/coal equivalent ratio	1.15	Reference ³⁷
Air flow rate, lb/hr (kg/hr)	3,965,800 (1,798,887)	Simulated, Reference ³⁷
Coal feed rate, lb/hr (kg/hr)	360,802 (163,660)	Simulated, Reference ³⁷
Steam generation		
Hot reheat steam, lb/hr (kg/hr)	3,027,000 (1,372,789)	Simulated
Superheat steam, lb/hr (kg/hr)	3,415,000 (1,548,753)	
HP steam generated from heat recovery in LSCC process, lb/hr (kg/hr)	2,976,000 (1,349,914)	
Steam condensate, lb/hr (kg/hr)	5,778,051 (2,620,924)	
Heat duty of cooling tower, mmBtu/hr (MJ/hr)	2,918 (3,078,800)	
Carbonator		
Inlet flue gas flow rate, lb/hr (kg/hr)	4,294,504 (1,947,987)	Simulated
Gas inlet temperature, °C	340	Simulated, Calculated
Gas inlet pressure, psia (atm)	14.5 (0.99)	Simulated
SO ₂ inlet, lb/hr (kg/hr)	16,473 (7,472)	Simulated
Sorbent/CO ₂ molar ratio	1.0	Assumed
CaO in, lb/hr (kg/hr)	1,066,793 (483,897)	Calculated
Operating temperature, °C	650	Reference ^{41,42,48,49}
Pressure drop, psi (atm)	1.47 (0.10)	Calculated
CO ₂ capture efficiency, wt %	90	Assumed
CO ₂ captured, lb/hr (kg/hr)	834,857 (378,691)	Simulated
SO ₂ capture efficiency, wt %	100	Assumed
SO ₂ captured, lb/hr (kg/hr)	16,473 (7,472)	Simulated
Calciner		
Sorbent make-up, lb/hr (kg/hr)	0	Assumed
CaCO ₃ feed in, lb/hr (kg/hr)	1,898,620 (861,214)	Calculated

Table 14 (cont.)

Parameter	PC plant with LSCC	Data source
CaCO ₃ temperature in (before mixing with CO ₂ /O ₂ /NG gas stream), °C	650	Reference ^{41,42,48,49}
O ₂ set-point in inlet flow of CO ₂ /O ₂ /NG gas stream, mol %	50	Assumed
O ₂ flow rate (95% purity), lb/hr (kg/hr)	353,700 (160,483)	Simulated
Natural gas flow rate, lb/hr (kg/hr)	83,500 (37,876)	Simulated
Percent of product gas recirculated	16	Assumed
Operating temperature, °C	950	Reference ⁴²
Pressure drop in reactor, psi (atm)	1.18 (0.08)	Calculated
<i>CO₂ product</i>		
Flow rate, lb/hr (kg/hr)	921,716 (418,090)	Simulated
Temperature, °C	20	Reference ^{45,48,49}
Pressure, psia (atm)	1,205 (82)	Reference ¹⁴

Appendix C3 Carbonator

1) Pressure Drop Calculation for Carbonator^a

$$\Delta P = \text{Density} \cdot (9.8 \text{ m/s}^2) \cdot \text{height}$$

$$\text{Density} = [(\text{Gas density}) \cdot (\text{voidage})] + [(\text{Solid density}) \cdot (1 - \text{voidage})]$$

$$\text{Density} = (0.1 \text{ kg/m}^3) \cdot 0.99 + (3,500 \text{ kg/m}^3) \cdot 0.01$$

$$\text{Density} = 35.1 \text{ kg/m}^3$$

$$\Delta P = (35.1 \text{ kg/m}^3) \cdot (9.8 \text{ m/s}^2) \cdot (30 \text{ m}), \text{ assuming a reactor height of 30 m.}$$

$$\Delta P = 0.10 \text{ atm (1.47 psi)}$$

^a Shimuzu used a voidage of 0.82 for a bubbling fluidized bed reactor; a voidage of 0.99 is assumed to be reasonable for the circulating fluidized bed reactor in this study^{42,55}

2) Calculation of Temperature for Post-Boiler Flue Gas Entering Carbonator^a

$$(C_{p,\text{Flue},650\text{C}}) \cdot (M_{\text{Flue}}) \cdot (\Delta T) = [(C_{p,\text{CaO},950\text{C}} + C_{p,\text{CaO},650\text{C}})/2] \cdot (M_{\text{CaO}}) \cdot (950 \text{ }^\circ\text{C} - 650 \text{ }^\circ\text{C})$$

$$(0.3 \text{ kJ/kg-}^\circ\text{C}) \cdot (541 \text{ kg/s}) \cdot (\Delta T) = [1.2 \text{ kJ/kg-}^\circ\text{C}] \cdot (134 \text{ kg/s}) \cdot (300 \text{ }^\circ\text{C})$$

$$\Delta T = 297 \text{ }^\circ\text{C}$$

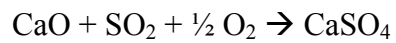
$$T_{\text{Flue,in}} = 650 \text{ }^\circ\text{C} - 297 \text{ }^\circ\text{C}$$

$$T_{\text{Flue,in}} = 353 \text{ }^\circ\text{C}$$

^a The flue gas temperature after the economizer and before the SCR is 340 °C – for practical purposes, this temperature is employed in the process simulation instead of the calculated 353 °C to provide the required flue gas/CaO mixture temperature within 10 °C of 650 °C

3) Loss of Available CaO Due to Reaction with Post-boiler Flue Gas SO₂

Reaction:



CaO consumed (see Table C.2)

$$= (\text{Mass flow rate of SO}_2 \text{ into carbonator}) * [(\text{CaO molar mass})/(\text{SO}_2 \text{ molar mass})]$$

$$= (16,473 \text{ lb/hr SO}_2) * [(56.1 \text{ grams/mol CaO})/(64.1 \text{ grams/mol SO}_2)]$$

$$\text{CaO consumed} = 14,417 \text{ lb/hr CaO (6,540 kg/hr)}$$

Percent of CaO into carbonator (see Table C.2)

$$= 100 * (\text{Mass flow rate of CaO consumed})/(\text{Mass flow rate of CaO into carbonator})$$

$$= 100 * (14,417 \text{ lb/hr CaO})/(1,066,793 \text{ lb/hr CaO})$$

$$1.4\% \text{ of CaO into carbonator consumed}$$

4) Pneumatic Transportation of CaO^{a,b,c}

Flue gas mass flow rate requirement

$$= (\text{Mass flow rate of CaO into carbonator}) * (\text{Required gas-to-solid ratio by weight})$$

$$= (1,066,793 \text{ lb/hr CaO}) * 0.1$$

$$= 106,679 \text{ lb/hr (48,380 kg/hr) from boiler flue gas required}$$

$$= (2.5 \text{ mbar/m}) * (100 \text{ m})$$

$$= 250 \text{ mbar}$$

$$= 3.6 \text{ psi (0.24 atm)}^a \text{ Gas-to-solid ratio assumed to be 0.1 by weight}^{56}$$

$$^b \text{ Pressure drop assumed to be 2.5 mbar/m}^{56}$$

$$^c \text{ Assume 100 m of ducting}$$

Appendix C4 Calcliner

1) Pressure Drop Calculation for Calcliner^a

$$\Delta P = \text{Density} \cdot (9.8 \text{ m/s}^2) \cdot \text{height}$$

$$\text{Density} = [(\text{Gas density}) \cdot (\text{voidage})] + [(\text{Solid density}) \cdot (1 - \text{voidage})]$$

$$\text{Density} = (0.1 \text{ kg/m}^3) \cdot 0.99 + (2,830 \text{ kg/m}^3) \cdot 0.01$$

$$\text{Density} = 28.4 \text{ kg/m}^3$$

$$\Delta P = (28.4 \text{ kg/m}^3) \cdot (9.8 \text{ m/s}^2) \cdot (30 \text{ m}), \text{ assuming a reactor height of 30 m.}$$

$$\Delta P = 0.08 \text{ atm (1.18 psi)}$$

^a Shimuzu used a voidage of 0.82 for a bubbling fluidized bed reactor; a voidage of 0.99 is assumed to be reasonable for the fluidized bed reactor in this study^{42,55}

2) Calculation for Natural Gas

The MW (thermal) value (Higher Heating Value) for natural gas, used in the calculation of net plant efficiency, was calculated by assuming 23,811 Btu/lb (55,391 kJ/kg) natural gas. The natural gas was assumed to consist of 100% methane.

3) Air Separation Unit Energy Use

Based on value of 160 kWh/short ton (176 kWh/metric ton) O₂ from Air Liquide.⁵⁰

4) Pneumatic Transportation of CaCO₃^{a,b,c}

Flue gas mass flow rate requirement

$$= (\text{Mass flow rate of CaCO}_3 \text{ into carbonator}) \cdot (\text{Required gas-to-solid ratio by weight})$$

$$= (1,898,620 \text{ lb/hr CaCO}_3) \cdot 0.1$$

$$= 189,862 \text{ lb/hr (86,121 kg/hr) gas stream required}$$

Use 206,058 lb/hr (93,468 kg/hr) from the CO₂-rich recycle gas stream from the ChemCAD simulation (see Table E.3).

$$= (2.5 \text{ mbar/m}) \cdot (100 \text{ m})$$

= 250 mbar

= 3.6 psi (0.24 atm)

^a Gas-to-solid ratio assumed to be 0.1 by weight⁵⁶

^b Pressure drop assumed to be 2.5 mbar/m⁵⁶

^c Assume 100 m of ducting

Appendix D Impacts of CO₂ Capture on PC Power Plants (650 MW_e)

Table 15 Impacts of CO₂ capture on PC power plant (650 MW_e gross)

	Illinois			PRB			70/30 PRB/IL blend			Illinois mine-mouth		
	<i>PC</i>	<i>PC-CC</i>	<i>Increase</i>	<i>PC</i>	<i>PC-CC</i>	<i>Increase</i>	<i>PC</i>	<i>PC-CC</i>	<i>Increase</i>	<i>PC</i>	<i>PC-CC</i>	<i>Increase</i>
Breakeven delivered Illinois coal cost against PRB coal cost of \$27.5/short ton (\$30.3/metric ton)												
Cost (\$/short ton)	33.0	36.0	3.0	-	-	-	39.5	41.0	1.5	21.0	18.5	-2.5
Cost (\$/metric ton)	36.4	39.7	3.3	-	-	-	43.5	45.2	1.7	23.1	20.4	-2.7
Atmospheric emissions (short tons/hour)												
CO ₂	6.08E+02	6.08E+01	-5.47E+02	6.34E+02	6.34E+01	-5.71E+02	6.21E+02	6.21E+01	-5.59E+02	5.96E+02	5.96E+01	-5.36E+02
SO ₂	7.20E-01	2.27E-03	-7.18E-01	7.41E-01	2.46E-03	-7.38E-01	7.32E-01	2.42E-03	-7.30E-01	7.18E-01	2.20E-03	-7.16E-01
NO _x	9.93E-02	9.75E-02	-1.80E-03	1.05E-01	1.03E-01	-2.00E-03	1.03E-01	1.01E-01	-2.00E-03	1.00E-01	9.83E-02	-1.70E-03
PM	8.70E-02	4.35E-02	-4.35E-02	9.87E-02	4.93E-02	-4.94E-02	9.72E-02	4.86E-02	-4.86E-02	4.34E-02	2.17E-02	-2.17E-02
Mercury	2.08E-06	2.08E-06	0	6.98E-06	6.98E-06	0	2.36E-06	2.36E-06	0	2.44E-06	2.44E-06	0
Atmospheric emissions (metric tons/hour)												
CO ₂	5.52E+02	5.52E+01	-4.96E+02	5.75E+02	5.75E+01	-5.18E+02	5.63E+02	5.63E+01	-5.07E+02	5.41E+02	5.41E+01	-4.86E+02
SO ₂	6.53E-01	2.06E-03	-6.51E-01	6.72E-01	2.23E-03	-6.70E-01	6.64E-01	2.20E-03	-6.62E-01	6.51E-01	2.00E-03	-6.50E-01
NO _x	9.01E-02	8.85E-02	-1.63E-03	9.53E-02	9.34E-02	-1.81E-03	9.34E-02	9.16E-02	-1.81E-03	9.07E-02	8.92E-02	-1.54E-03
PM	7.89E-02	3.95E-02	-3.95E-02	8.95E-02	4.47E-02	-4.48E-02	8.82E-02	4.41E-02	-4.41E-02	3.94E-02	1.97E-02	-1.97E-02
Mercury	1.89E-06	1.89E-06	0	6.33E-06	6.33E-06	0	2.14E-06	2.14E-06	0	2.21E-06	2.21E-06	0
Atmospheric emissions (lb/MWh)												
CO ₂	2.06E+03	2.84E+02	-1.77E+03	2.11E+03	2.93E+02	-1.81E+03	2.06E+03	2.84E+02	-1.78E+03	2.02E+03	2.78E+02	-1.74E+03
SO ₂	2.43E+00	1.07E-02	-2.43E+00	2.47E+00	1.14E-02	-2.45E+00	2.43E+00	1.11E-02	-2.43E+00	2.43E+00	1.03E-02	-2.40E+00
NO _x	3.35E-01	4.59E-01	1.23E-01	3.48E-01	4.78E-01	1.28E-01	3.42E-01	4.63E-01	1.22E-01	3.37E-01	4.59E-01	1.21E-01
PM	2.93E-01	2.03E-01	-9.08E-02	3.29E-01	2.29E-01	-9.94E-02	3.24E-01	2.23E-01	-9.99E-02	1.46E-01	1.01E-01	-4.50E-02
Mercury	7.03E-06	9.72E-06	2.69E-06	2.32E-05	3.24E-05	9.15E-06	7.85E-06	1.08E-05	3.00E-06	8.20E-06	9.72E-06	1.51E-06

Table 15 (cont.)

Atmospheric emissions (kg/MWh)												
CO ₂	9.32E+02	1.29E+02	-8.03E+02	9.56E+02	1.33E+02	-8.23E+02	9.36E+02	1.29E+02	-8.07E+02	9.17E+02	1.26E+02	-7.91E+02
SO ₂	1.10E+00	4.85E-03	-1.10E+00	1.12E+00	5.16E-03	-1.11E+00	1.10E+00	5.05E-03	-1.10E+00	1.10E+00	4.65E-03	-1.09E+00
NO _x	1.52E-01	2.08E-01	5.58E-02	1.58E-01	2.17E-01	5.82E-02	1.55E-01	2.10E-01	5.52E-02	1.53E-01	2.08E-01	5.51E-02
PM	1.33E-01	9.20E-02	-4.12E-02	1.49E-01	1.04E-01	-4.51E-02	1.47E-01	1.01E-01	-4.53E-02	6.62E-02	4.58E-02	-2.04E-02
Mercury	3.19E-06	4.41E-06	1.22E-06	1.05E-05	1.47E-05	4.15E-06	3.56E-06	4.92E-06	1.36E-06	3.72E-06	4.41E-06	6.85E-07
Capital required (\$/kW-net)												
Total	1.92E+03	3.27E+03	1.36E+03	1.93E+03	3.35E+03	1.42E+03	1.92E+03	3.30E+03	1.38E+03	2.00E+03	3.55E+03	1.54E+03
Boiler	1.14E+03	1.58E+03	4.35E+02	1.15E+03	1.61E+03	4.55E+02	1.14E+03	1.58E+03	4.37E+02	1.18E+03	1.62E+03	4.41E+02
CO ₂	0	6.45E+02	6.45E+02	0	6.58E+02	6.58E+02	0	6.49E+02	6.49E+02	0	8.20E+02	8.20E+02
SO ₂	4.43E+02	5.92E+02	1.49E+02	2.30E+02	3.22E+02	9.15E+01	2.50E+02	3.40E+02	8.99E+01	4.76E+02	6.31E+02	1.55E+02
NO _x	2.81E+02	3.88E+02	1.07E+02	2.87E+02	4.00E+02	1.13E+02	2.83E+02	3.91E+02	1.08E+02	2.79E+02	3.84E+02	1.04E+02
PM	5.01E+01	6.92E+01	1.91E+01	2.46E+02	3.43E+02	9.70E+01	2.44E+02	3.38E+02	9.34E+01	6.92E+01	9.50E+01	2.58E+01
Mercury	0	0	0	1.04E+01	1.44E+01	4.00E+00	0	0	0	0	0	0
Fixed O&M (\$/MWh)												
Total	8.55E+00	1.43E+01	5.78E+00	7.41E+00	1.28E+01	5.41E+00	7.47E+00	1.27E+01	5.27E+00	8.94E+00	1.54E+01	6.46E+00
Boiler	4.29E+00	5.92E+00	1.63E+00	4.29E+00	5.98E+00	1.69E+00	4.26E+00	5.89E+00	1.63E+00	4.45E+00	6.11E+00	1.66E+00
CO ₂	0	2.44E+00	2.44E+00	0	2.48E+00	2.48E+00	0	2.44E+00	2.44E+00	0	3.05E+00	3.05E+00
SO ₂	3.33E+00	4.67E+00	1.34E+00	1.88E+00	2.63E+00	7.50E-01	1.98E+00	2.71E+00	7.27E-01	3.53E+00	4.92E+00	1.39E+00
NO _x	7.31E-01	1.01E+00	2.79E-01	7.46E-01	1.04E+00	2.94E-01	7.36E-01	1.02E+00	2.84E-01	7.26E-01	9.97E-01	2.71E-01
PM	2.05E-01	2.84E-01	7.90E-02	4.80E-01	6.68E-01	1.88E-01	4.82E-01	6.67E-01	1.85E-01	2.34E-01	3.21E-01	8.70E-02
Mercury	0	0	0.00E+00	1.60E-02	2.20E-02	6.00E-03	0	0	0	0	0	0
Variable O&M (\$/MWh)												
Total	1.95E+01	4.88E+01	2.93E+01	1.96E+01	4.94E+01	2.98E+01	1.89E+01	4.92E+01	3.02E+01	1.85E+01	4.37E+01	2.52E+01
Boiler	1.51E+01	2.65E+00	-1.25E+01	1.65E+01	3.38E+00	-1.31E+01	1.62E+01	3.60E+00	-1.26E+01	1.23E+01	-2.06E-01	-1.25E+01
CO ₂	0	3.99E+01	3.99E+01	0	4.16E+01	4.16E+01	0	4.20E+01	4.20E+01	0	3.87E+01	3.87E+01

Table 15 (cont.)

SO ₂	2.82E+00	4.13E+00	1.31E+00	4.78E-01	7.17E-01	2.39E-01	1.31E+00	1.61E+00	2.95E-01	3.85E+00	5.52E+00	1.68E+00
NO _x	9.06E-01	1.25E+00	3.46E-01	9.46E-01	1.32E+00	3.73E-01	9.26E-01	1.28E+00	3.54E-01	8.88E-01	1.22E+00	3.32E-01
PM	6.20E-01	8.56E-01	2.36E-01	4.74E-01	6.60E-01	1.87E-01	5.30E-01	7.32E-01	2.03E-01	1.42E+00	1.95E+00	5.31E-01
Mercury	0	0	0	1.19E+00	1.66E+00	4.69E-01	0	0	0	0	0	0

Table 16 Impacts of CO₂ capture on PC power plant (175 MW_e gross)

	Illinois			PRB			70/30 PRB/IL blend			Illinois mine-mouth		
	PC	PC-CC	Increase	PC	PC-CC	Increase	PC	PC-CC	Increase	PC	PC-CC	Increase
Breakeven delivered Illinois coal cost against PRB coal cost of \$27.5/short ton (\$30.3/metric ton)												
Cost (\$/short ton)	21.0	21.5	0.5	-	-	-	36.0	39.5		6.5	1.5	-5.0
Cost (\$/metric ton)	23.1	23.7	0.6	-	-	-	39.7	43.5	0.0	7.2	1.7	-5.5
Atmospheric emissions (short tons/hour)												
CO ₂	1.64E+02	1.64E+01	-1.48E+02	1.71E+02	1.71E+01	-1.54E+02	1.67E+02	1.67E+01	-1.50E+02	1.61E+02	1.61E+01	-1.44E+02
SO ₂	1.94E-01	6.10E-04	-1.93E-01	2.00E-01	6.60E-04	-1.99E-01	1.97E-01	6.50E-04	-1.97E-01	1.93E-01	5.90E-04	-1.93E-01
NO _x	2.26E-02	2.22E-02	-4.00E-04	2.39E-02	2.34E-02	-5.00E-04	2.34E-02	2.30E-02	-4.00E-04	2.28E-02	2.24E-02	-4.00E-04
PM	2.34E-02	1.17E-02	-1.17E-02	2.66E-02	1.00E-05	-2.66E-02	2.62E-02	1.31E-02	-1.31E-02	1.17E-02	5.84E-03	-5.84E-03
Mercury	5.61E-07	5.61E-07	0	1.88E-06	1.88E-06	-2.00E-09	8.61E-07	6.36E-07	-2.25E-07	6.56E-07	6.56E-07	0
Atmospheric emissions (metric tons/hour)												
CO ₂	1.49E+02	1.49E+01	-1.34E+02	1.55E+02	1.55E+01	-1.40E+02	1.52E+02	1.52E+01	-1.36E+02	1.46E+02	1.46E+01	-1.31E+02
SO ₂	1.76E-01	5.53E-04	-1.75E-01	1.81E-01	5.99E-04	-1.81E-01	1.79E-01	5.90E-04	-1.79E-01	1.75E-01	5.35E-04	-1.75E-01
NO _x	2.05E-02	2.01E-02	-3.63E-04	2.17E-02	2.12E-02	-4.54E-04	2.12E-02	2.09E-02	-3.63E-04	2.07E-02	2.03E-02	-3.63E-04
PM	2.12E-02	1.06E-02	-1.06E-02	2.41E-02	9.07E-06	-2.41E-02	2.38E-02	1.19E-02	-1.19E-02	1.06E-02	5.30E-03	-5.30E-03
Mercury	5.09E-07	5.09E-07	0	1.71E-06	1.71E-06	-1.81E-09	7.81E-07	5.77E-07	-2.04E-07	5.95E-07	5.95E-07	0
Atmospheric emissions (lb/MWh)												
CO ₂	2.84E+03	2.84E+02	-2.56E+03	2.91E+03	2.91E+02	-2.62E+03	2.84E+03	2.84E+02	-2.56E+03	2.78E+03	2.78E+02	-2.51E+03
SO ₂	2.43E+00	1.06E-02	-2.43E+00	2.45E+00	1.13E-02	-2.43E+00	2.43E+00	1.11E-02	-2.40E+00	2.43E+00	1.03E-02	-2.43E+00
NO _x	2.82E-01	3.84E-01	1.02E-01	2.93E-01	3.99E-01	1.07E-01	2.87E-01	3.90E-01	1.03E-01	2.87E-01	3.88E-01	1.01E-01
PM	2.93E-01	2.03E-01	-9.02E-02	3.26E-01	1.10E-04	-3.26E-01	3.20E-01	2.23E-01	-9.90E-02	1.47E-01	1.01E-01	-4.59E-02
Mercury	7.01E-06	9.70E-06	2.69E-06	2.32E-05	3.22E-05	9.06E-06	1.05E-05	1.08E-05	2.29E-07	8.27E-06	1.14E-05	3.13E-06

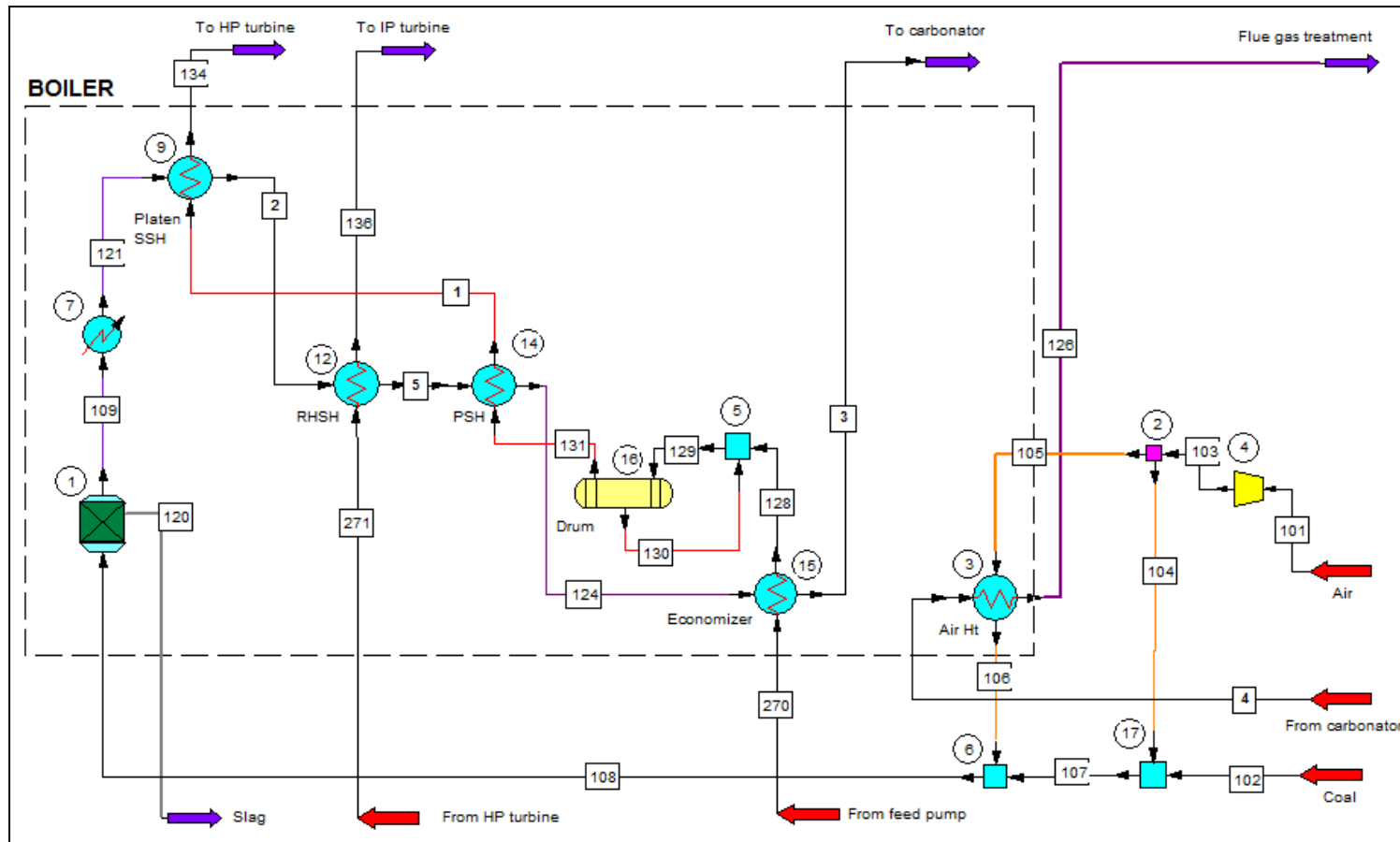
Table 16 (cont.)

Atmospheric emissions (kg/MWh)												
CO ₂	1.29E+03	1.29E+02	-1.16E+03	1.32E+03	1.32E+02	-1.19E+03	1.29E+03	1.29E+02	-1.16E+03	1.26E+03	1.26E+02	-1.14E+03
SO ₂	1.10E+00	4.81E-03	-1.10E+00	1.11E+00	5.13E-03	-1.10E+00	1.10E+00	5.02E-03	-1.09E+00	1.10E+00	4.66E-03	-1.10E+00
NO _x	1.28E-01	1.74E-01	4.61E-02	1.33E-01	1.81E-01	4.85E-02	1.30E-01	1.77E-01	4.68E-02	1.30E-01	1.76E-01	4.60E-02
PM	1.33E-01	9.20E-02	-4.09E-02	1.48E-01	5.00E-05	-1.48E-01	1.45E-01	1.01E-01	-4.49E-02	6.67E-02	4.59E-02	-2.08E-02
Mercury	3.18E-06	4.40E-06	1.22E-06	1.05E-05	1.46E-05	4.11E-06	4.78E-06	4.88E-06	1.04E-07	3.75E-06	5.16E-06	1.42E-06
Capital required (\$/kW-net)												
Total	2.92E+03	4.80E+03	1.88E+03	2.76E+03	4.59E+03	1.83E+03	2.76E+03	4.54E+03	1.78E+03	3.08E+03	5.19E+03	2.11E+03
Boiler	1.54E+03	2.13E+03	5.90E+02	1.54E+03	2.15E+03	6.09E+02	1.53E+03	2.11E+03	5.85E+02	1.60E+03	2.21E+03	6.05E+02
CO ₂	0	7.35E+02	7.35E+02	0	7.44E+02	7.44E+02	0	7.34E+02	7.34E+02	0	9.36E+02	9.36E+02
SO ₂	8.83E+02	1.24E+03	3.59E+02	4.15E+02	5.80E+02	1.65E+02	4.59E+02	6.23E+02	1.64E+02	9.57E+02	1.34E+03	3.79E+02
NO _x	4.30E+02	5.95E+02	1.65E+02	4.35E+02	6.07E+02	1.72E+02	4.30E+02	5.94E+02	1.64E+02	4.28E+02	5.89E+02	1.61E+02
PM	7.08E+01	9.81E+01	2.72E+01	3.49E+02	4.87E+02	1.38E+02	3.47E+02	4.79E+02	1.33E+02	9.45E+01	1.30E+02	3.57E+01
Mercury	0	0	0	1.58E+01	2.18E+01	6.07E+00	0	0	0	0	0	0
Fixed O&M (\$/MWh)												
Total	1.93E+01	3.03E+01	1.10E+01	1.62E+01	2.60E+01	9.85E+00	1.63E+01	2.59E+01	9.59E+00	2.01E+01	3.20E+01	1.19E+01
Boiler	9.57E+00	1.33E+01	3.68E+00	9.45E+00	1.32E+01	3.72E+00	9.42E+00	1.30E+01	3.60E+00	9.90E+00	1.36E+01	3.74E+00
CO ₂	0.00E+00	3.44E+00	3.44E+00	0.00E+00	3.46E+00	3.46E+00	0.00E+00	3.42E+00	3.42E+00	0.00E+00	4.15E+00	4.15E+00
SO ₂	7.91E+00	1.11E+01	3.20E+00	4.25E+00	5.94E+00	1.69E+00	4.49E+00	6.14E+00	1.65E+00	8.39E+00	1.17E+01	3.31E+00
NO _x	1.27E+00	1.76E+00	4.92E-01	1.28E+00	1.79E+00	5.08E-01	1.27E+00	1.75E+00	4.80E-01	1.26E+00	1.74E+00	4.80E-01
PM	5.36E-01	7.42E-01	2.06E-01	1.14E+00	1.58E+00	4.40E-01	1.14E+00	1.58E+00	4.40E-01	5.80E-01	7.98E-01	2.18E-01
Mercury	0	0	0	5.80E-02	8.10E-02	2.30E-02	0	0	0	0	0	0
Variable O&M (\$/MWh)												
Total	1.95E+01	5.34E+01	3.38E+01	1.95E+01	5.34E+01	3.39E+01	1.89E+01	5.34E+01	3.44E+01	1.85E+01	4.83E+01	2.98E+01
Boiler	1.45E+01	-3.80E+00	-1.83E+01	1.62E+01	-2.65E+00	-1.89E+01	1.60E+01	-2.23E+00	-1.82E+01	1.15E+01	-7.08E+00	-1.86E+01
CO ₂	0	4.99E+01	4.99E+01	0	5.14E+01	5.14E+01	0	5.17E+01	5.17E+01	0	4.89E+01	4.89E+01

Table 16 (cont.)

SO ₂	3.18E+00	4.70E+00	1.52E+00	5.73E-01	8.50E-01	2.77E-01	1.42E+00	1.76E+00	3.37E-01	4.30E+00	6.22E+00	1.93E+00
NO _x	9.40E-01	1.30E+00	3.61E-01	9.77E-01	1.36E+00	3.86E-01	9.56E-01	1.32E+00	3.66E-01	9.26E-01	1.28E+00	3.49E-01
PM	9.09E-01	1.26E+00	3.50E-01	5.22E-01	7.29E-01	2.06E-01	5.78E-01	7.99E-01	2.21E-01	1.83E+00	2.51E+00	6.88E-01
Mercury	0	0	0	1.19E+00	1.66E+00	4.70E-01	0	0	0	0	0	0

Appendix E Power Plant with 533 MW_e (Gross) Output Installed with the LSCC Process – Simulation Results of Mass and Energy Balance Calculations from ChemCAD



Note: SSH = Superheater; RHSH = Reheat Superheater; PSH = Pre-superheater

Figure 23 Boiler

Table 17 Boiler stream conditions

Stream number	1	2	3	4	5	101	102	103
Mass flow, kg/hr	1,549,018	1,947,993	1,947,993	1,569,269	1,947,993	1,798,857	163,657	1,798,857
Mass flow, lb/hr	89,943	89,943	89,943	89,943	89,943	89,943	89,943	89,943
Temperature, °C	379.3	896.0	340.0	339.5	619.4	17.2	19.4	26.4
Pressure, atm	166.03	1.03	0.99	0.85	1.01	0.98	1.08	1.08
Pressure, psia	2440.0	15.1	14.5	12.5	14.8	14.4	15.9	15.9
Vapor mass fraction	1.000	1.000	1.000	1.000	1.000	1.000	0	1.000
Enthalpy, MJ/hr	-2.040E+07	-2.926E+06	-4.217E+06	-9.473E+05	-3.588E+06	-1.462E+04	-3.516E+05	2.026E+03
Enthalpy, mmBtu/hr	-1.933E+04	-2.773E+03	-3.997E+03	-8.979E+02	-3.401E+03	-1.386E+01	-3.333E+02	1.921E+00
Stream number	104	105	106	107	108	109	120	121
Mass flow, kg/hr	89,943	1,708,914	1,708,914	253,600	1,962,505	1,947,993	14,558	1,947,993
Mass flow, lb/hr	198,324	3,768,155	3,768,155	559,188	4,327,325	4,295,324	32,099	4,295,324
Temperature, °C	26.4	26.4	217.7	12.5	189.2	1,315.6	1,315.6	1,251.7
Pressure, atm	1.08	1.08	1.08	1.08	1.08	1.04	1.04	1.04
Pressure, psia	15.9	15.9	15.9	15.9	15.9	15.3	15.3	15.3
Vapor mass fraction	1.000	1.000	1.000	0.358	0.922	1.000	0	1.000
Enthalpy, MJ/hr	1.013E+02	1.925E+03	3.359E+05	-3.515E+05	-1.563E+04	-1.866E+06	-4.208E+04	-2.030E+06
Enthalpy, mmBtu/hr	9.603E-02	1.825E+00	3.184E+02	-3.332E+02	-1.481E+01	-1.768E+03	-3.988E+01	-1.924E+03
Stream number	124	126	128	129	130	131	134	136
Mass flow, kg/hr	1,947,993	1,569,269	1,549,018	1,549,018	0	1,549,018	1,550,044	1,370,411
Mass flow, lb/hr	4,295,324	3,460,239	3,415,584	3,415,584	0	3,415,584	3,417,847	3,021,756
Temperature, °C	482.0	146.1	294.5	294.5	0	351.7	538.0	538.0
Pressure, atm	1.00	0.83	190.53	166.71	0	166.71	164.31	37.11
Pressure, psia	14.7	12.2	2800.0	2450.0	0	2450.0	2414.7	545.4
Vapor mass fraction	1.000	1.000	0	0	0	1.000	1.000	1.000
Enthalpy, MJ/hr	-3.903E+06	-1.281E+06	-2.271E+07	-2.271E+07	0	-2.071E+07	-1.950E+07	-1.706E+07
Enthalpy, mmBtu/hr	-3.699E+03	-1.214E+03	-2.152E+04	-2.152E+04	0	-1.963E+04	-1.849E+04	-1.617E+04
Stream number	270	271						
Mass flow, kg/hr	1,549,018	1,370,411						
Mass flow, lb/hr	3,415,584	3,021,756						
Temperature, °C	255.69	333.90						
Pressure, atm	190.53	41.07						
Pressure, psia	2800.0	603.6						
Vapor mass fraction	0	1.000						
Enthalpy, MJ/hr	-2.302E+07	-1.773E+07						
Enthalpy, mmBtu/hr	-2.182E+04	-1.680E+04						

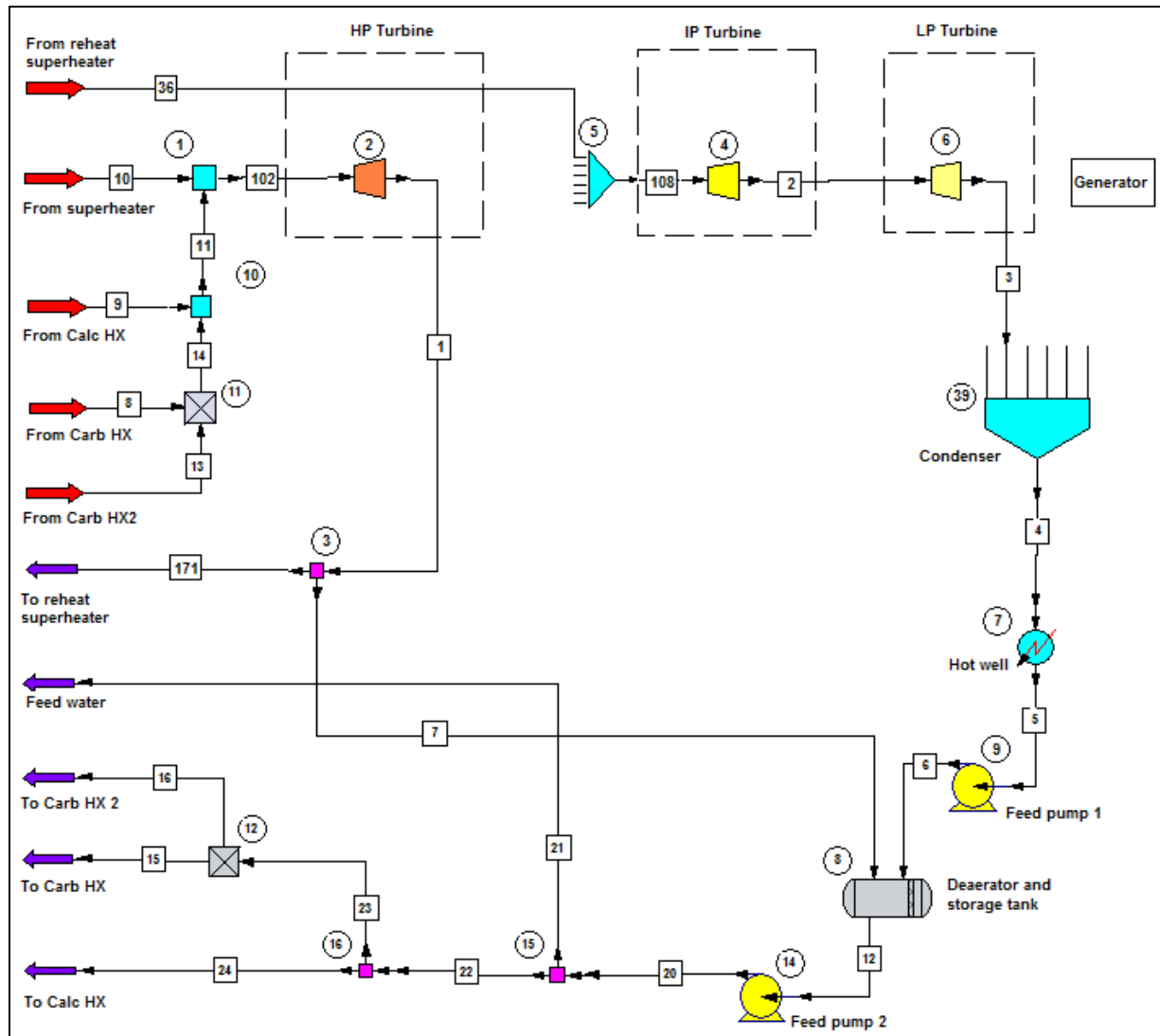


Figure 24 Steam cycle

Table 18 Stream conditions for steam cycle

Stream number	1	2	3	4	5	6	7	8
Mass flow, kg/hr	2,898,909	1,373,024	1,373,024	1,373,024	1,373,024	1,373,024	1,525,885	836,878
Mass flow, lb/hr	6,392,094	3,027,518	3,027,518	3,027,518	3,027,518	3,027,518	3,364,576	1,845,316
Temperature, °C	333.9	375.2	42.1	36.7	35.8	36.9	333.9	538.0
Pressure, atm	41.10	11.90	0.10	0.10	0.10	40.00	41.10	164.30
Pressure, psia	604.0	174.9	1.5	1.5	1.5	587.8	604.0	2,414.6
Vapor mass fraction	1.000	1.000	0.923	0.927	0	0	1.000	1.000
Enthalpy, MJ/hr	-3.749E+07	-1.754E+07	-1.866E+07	-1.866E+07	-2.174E+07	-2.173E+07	-1.974E+07	-1.053E+07
Enthalpy, mmBtu/hr	-3.554E+04	-1.663E+04	-1.769E+04	-1.769E+04	-2.060E+04	-2.060E+04	-1.871E+04	-9.981E+03
Stream number	9	10	11	12	13	14	15	16
Mass flow, kg/hr	265,352	1,549,018	1,349,891	2,898,909	247,661	1,084,539	836,878	247,661
Mass flow, lb/hr	585,100	3,415,585	2,976,509	6,392,093	546,093	2,391,409	1,845,316	546,093
Temperature, °C	538.0	538.0	538.0	251.1	538.0	538.0	255.7	255.7
Pressure, atm	164.30	164.30	164.30	40.00	164.30	164.30	190.50	190.50
Pressure, psia	2,414.6	2,414.6	2,414.6	587.8	2,414.6	2,414.6	2,799.6	2,799.6
Vapor mass fraction	1.000	1.000	1.000	0	1.000	1.000	0	0
Enthalpy, MJ/hr	-3.339E+06	-1.949E+07	-1.699E+07	-4.315E+07	-3.116E+06	-1.365E+07	-1.244E+07	-3.681E+06
Enthalpy, mmBtu/hr	-3.165E+03	-1.847E+04	-1.610E+04	-4.090E+04	-2.954E+03	-1.293E+04	-1.179E+04	-3.488E+03
Stream number	20	21	22	23	24	36	102	108
Mass flow, kg/hr	2,898,909	1,549,018	1,349,891	1,084,539	265,352	1,373,024	2,898,909	1,373,024
Mass flow, lb/hr	6,392,093	3,415,584	2,976,509	2,391,409	585,100	3,027,518	6,392,094	3,027,518
Temperature, °C	255.7	255.7	255.7	255.7	255.7	538.0	538.0	538.0
Pressure, atm	190.50	190.50	190.50	190.50	190.50	37.10	164.30	37.10
Pressure, psia	2,799.6	2,799.6	2,799.6	2,799.6	2,799.6	545.2	2,414.6	545.2
Vapor mass fraction	0	0	0	0	0	1.000	1.000	1.000
Enthalpy, MJ/hr	-4.308E+07	-2.302E+07	-2.006E+07	-1.612E+07	-3.943E+06	-1.709E+07	-3.648E+07	-1.709E+07
Enthalpy, mmBtu/hr	-4.083E+04	-2.182E+04	-1.901E+04	-1.528E+04	-3.738E+03	-1.620E+04	-3.457E+04	-1.620E+04
Stream number	171							
Mass flow, kg/hr	1,373,024							
Mass flow, lb/hr	3,027,518							
Temperature, °C	333.9							
Pressure, atm	41.10							
Pressure, psia	604.0							
Vapor mass fraction	1.000							
Enthalpy, MJ/hr	-1.776E+07							
Enthalpy, mmBtu/hr	-1.683E+04							

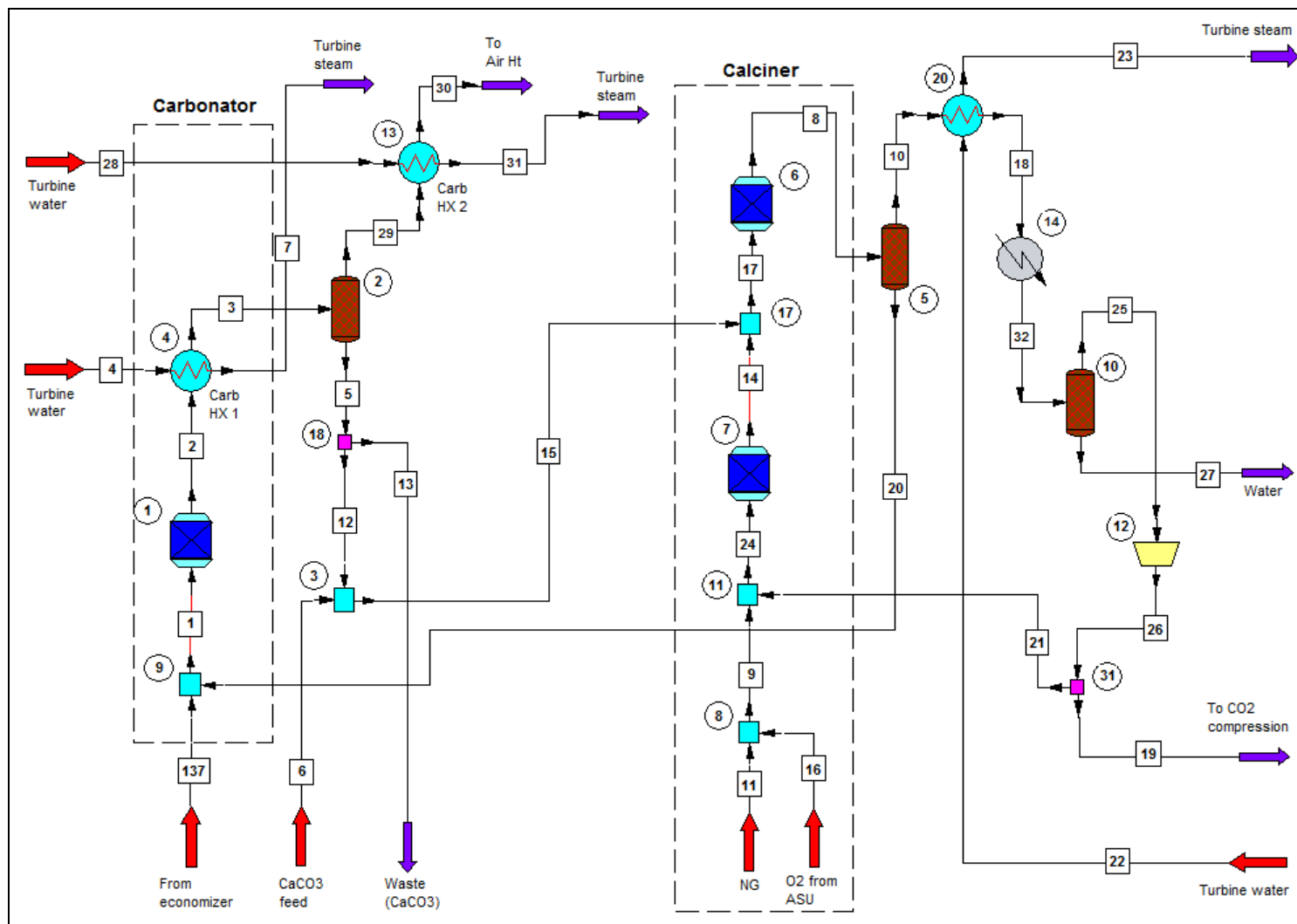
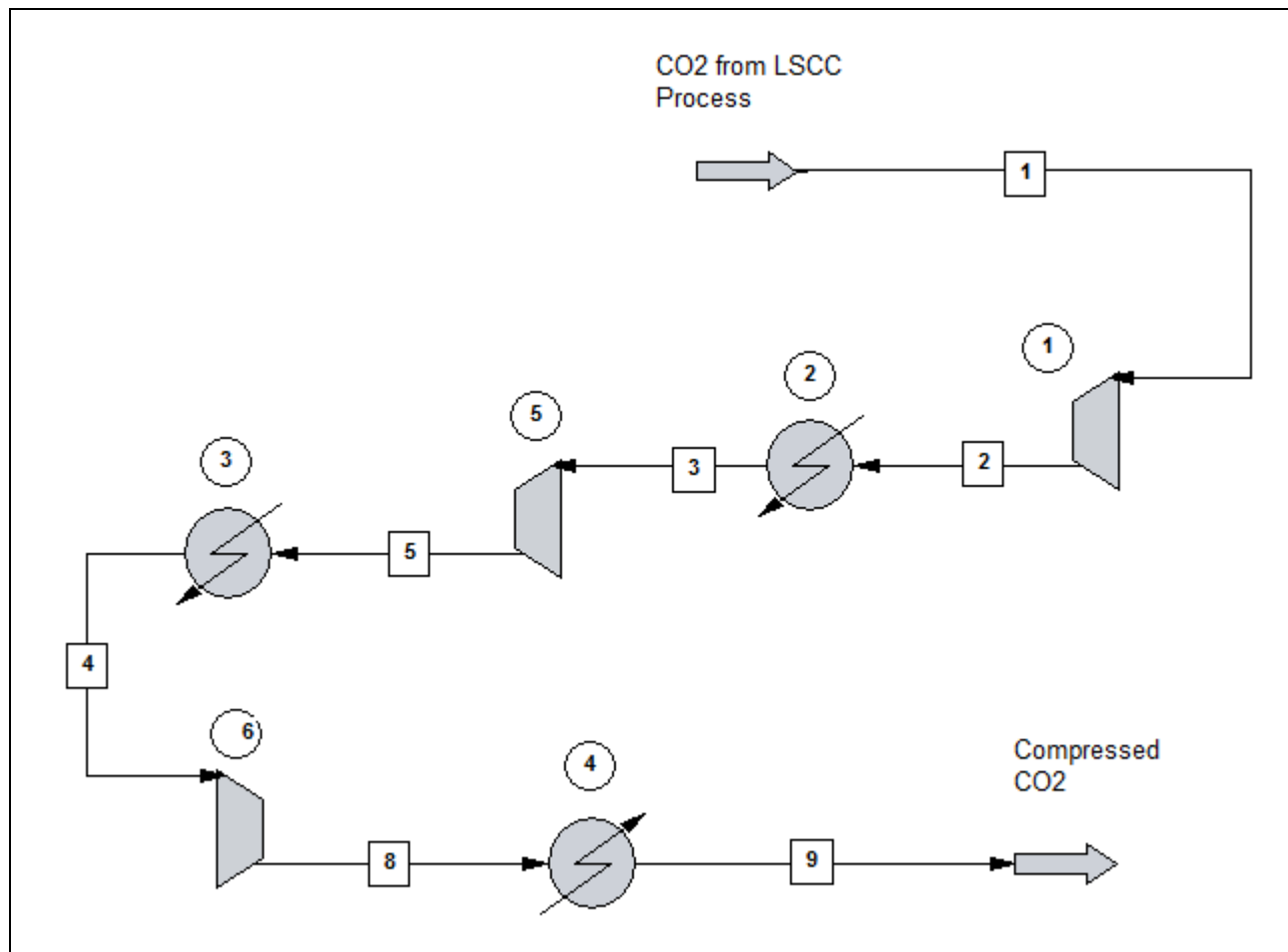


Figure 25 LSCC process

Table 19 Stream conditions for LSCC process

Stream number	1	2	3	4	5	6	7	8
Mass flow, kg/hr	2,430,470	2,430,470	2,430,470	836,878	861,201	0	836,878	1,152,959
Mass flow, lb/hr	5,359,186	5,359,186	5,359,186	1,845,316	1,898,948	0	1,845,316	2,542,275
Temperature, °C	634.9	1,165.5	650.0	255.0	650.0	0	538.0	950.0
Pressure, atm	0.99	0.89	0.87	190.53	0.86	0	164.31	1.00
Pressure, psia	14.5	13.1	12.8	2,800.0	12.6	0	2,414.7	14.7
Vapor mass fraction	0.801	0.646	0.646	0	0	0	1.000	0.581
Enthalpy, MJ/hr	-1.357E+07	-1.357E+07	-1.548E+07	-1.244E+07	-1.510E+07	0	-1.053E+07	-1.486E+07
Enthalpy, mmBtu/hr	-1.286E+04	-1.286E+04	-1.468E+04	-1.179E+04	-1.431E+04	0	-9.981E+03	-1.409E+04
Stream number	9	10	11	12	13	14	15	16
Mass flow, kg/hr	198,311	670,444	37,875	861,201	0	291,759	861,201	160,436
Mass flow, lb/hr	437,275	1,478,329	83,514	1,898,948	0	643,328	1,898,948	353,761
Temperature, °C	29.2	950.0	27.0	650.0	0	3,586.8	649.9	30.5
Pressure, atm	1.08	0.98	1.08	0.86	0	1.08	0.96	1.08
Pressure, psia	15.9	14.4	15.9	12.6	0	15.9	14.1	15.9
Vapor mass fraction	1.000	1.000	1.000	0	0	1.000	0	1.000
Enthalpy, MJ/hr	-1.758E+05	-5.505E+06	-1.766E+05	-1.510E+07	0	-9.911E+05	-1.510E+07	7.781E+02
Enthalpy, mmBtu/hr	-1.666E+02	-5.217E+03	-1.674E+02	-1.431E+04	0	-9.394E+02	-1.431E+04	7.375E-01
Stream number	17	18	19	20	21	22	23	24
Mass flow, kg/hr	1,152,959	670,444	490,614	482,516	93,450	265,352	265,352	291,761
Mass flow, lb/hr	2,542,275	1,478,329	1,081,804	1,063,948	206,058	585,100	585,100	643,333
Temperature, °C	1,408.6	260.5	104.8	950.0	104.8	255.0	538.0	49.3
Pressure, atm	1.08	0.97	1.08	0.98	1.08	190.53	164.31	1.08
Pressure, psia	15.9	14.3	15.9	14.4	15.9	2,800.0	2,414.7	15.9
Vapor mass fraction	0.253	1.000	1.000	0	1.000	0	1.000	1.000
Enthalpy, MJ/hr	-1.609E+07	-6.110E+06	-4.281E+06	-9.358E+06	-8.153E+05	-3.944E+06	-3.339E+06	-9.911E+05
Enthalpy, mmBtu/hr	-1.525E+04	-5.791E+03	-4.057E+03	-8.869E+03	-7.728E+02	-3.739E+03	-3.165E+03	-9.394E+02
Stream number	25	26	27	28	29	30	31	32
Mass flow, kg/hr	584,064	584,064	86,379	247,661	1,569,269	1,569,269	247,661	670,444
Mass flow, lb/hr	1,287,862	1,287,862	190,466	546,093	3,460,238	3,460,238	546,093	1,478,329
Temperature, °C	94.0	104.8	94.0	255.0	650.0	340.3	538.0	94.0
Pressure, atm	0.95	1.08	0.95	190.53	0.86	0.85	164.31	0.97
Pressure, psia	14.0	15.9	14.0	2,800.0	12.6	12.5	2,414.7	14.3
Vapor mass fraction	1.000	1.000	0.061	0	1.000	1.000	1.000	1.000
Enthalpy, MJ/hr	-5.102E+06	-5.096E+06	-1.317E+06	-3.681E+06	-3.809E+05	-9.460E+05	-3.116E+06	-6.234E+06
Enthalpy, mmBtu/hr	-4.835E+03	-4.830E+03	-1.248E+03	-3.489E+03	-3.610E+02	-8.966E+02	-2.954E+03	-5.908E+03
Stream number	137							
Mass flow, kg/hr	1,947,954							
Mass flow, lb/hr	4,295,239							
Temperature, °C	340.0							
Pressure, atm	0.99							
Pressure, psia	14.5							
Vapor mass fraction	1.000							
Enthalpy, MJ/hr	-4.216E+06							
Enthalpy, mmBtu/hr	-3.996E+03							



Note: Compressor efficiency assumed at 85%

Figure 26 CO₂ compression

Table 20 Stream conditions for CO₂ compression

Stream number	1	2	3	4	5	8	9
Mass flow, kg/hr	490,614	490,614	490,614	490,614	490,614	490,614	490,614
Mass flow, lb/hr	1,081,804	1,081,804	1,081,804	1,081,804	1,081,804	1,081,804	1,081,804
Temperature, °C	104.8	340.7	40.0	40.0	209.5	68.5	20.0
Pressure, atm	1.08	9.87	9.87	59.22	59.22	81.91	81.91
Pressure, psia	15.8	145.0	145.0	870.2	870.2	1,203.8	1,203.8
Vapor mass fraction	1.000	1.000	1.000	1.000	1.000	1.000	0
Enthalpy, MJ/hr	-4.281E+06	-4.164E+06	-4.313E+06	-4.340E+06	-4.241E+06	-4.332E+06	-4.434E+06
Enthalpy, mmBtu/hr	-4.057E+03	-3.947E+03	-4.088E+03	-4.114E+03	-4.019E+03	-4.106E+03	-4.202E+03

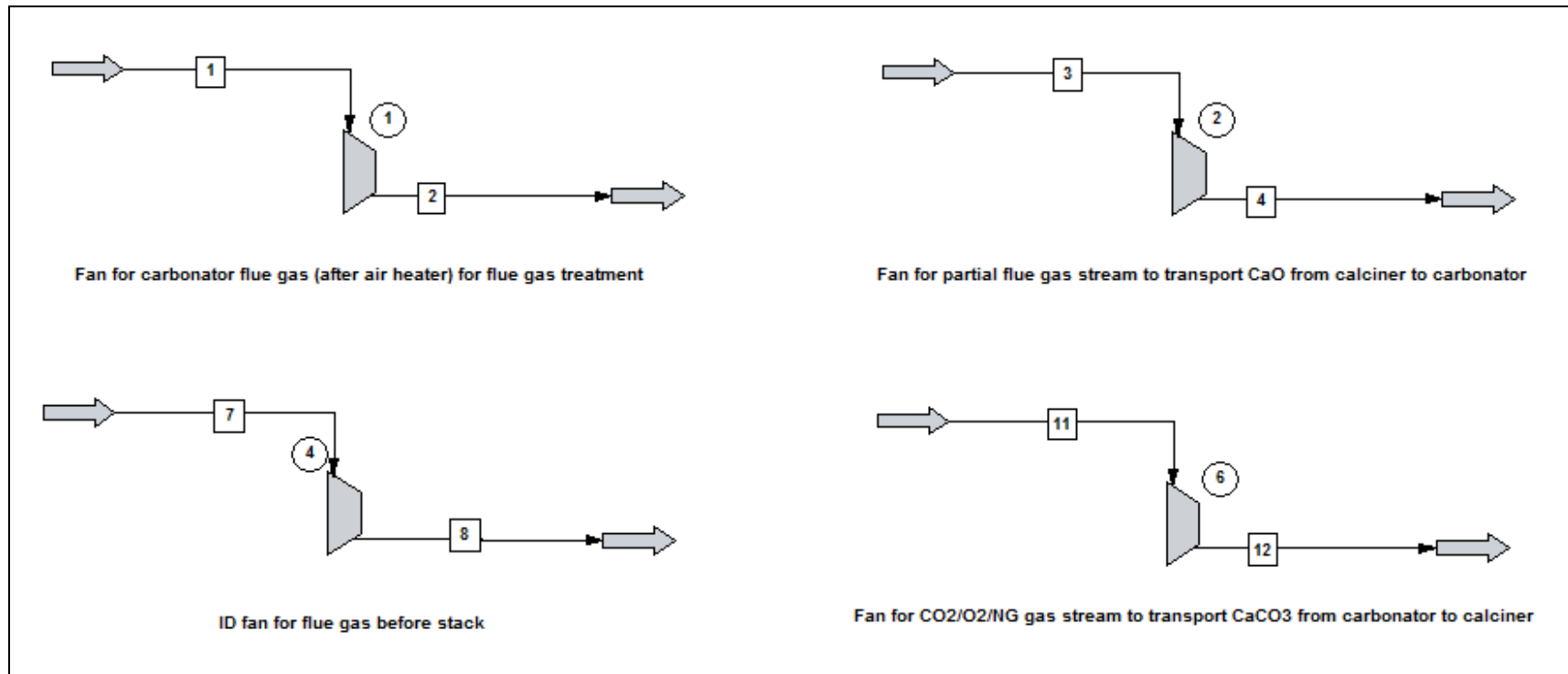


Figure 27 Compression of different gas streams

Table 21 Stream conditions for compression of different gas streams

Stream number	1	2	3	4	7	8	11	12
Mass flow, kg/hr	1,569,269	1,569,269	48,252	48,252	1,569,269	50,213	93,450	93,450
Mass flow, lb/hr	3,460,239	3,460,239	106,396	106,396	3,460,239	110,719	206,058	206,058
Temperature, °C	146.1	167.9	340.0	381.0	167.9	30.5	104.8	123.8
Pressure, atm	0.83	0.97	0.99	1.23	0.97	1.08	1.08	1.32
Pressure, psia	12.2	14.3	14.5	18.1	14.3	15.8	15.8	19.4
Vapor mass fraction	1.000	1.000	1.000	1.000	1.000	1.000	1.000	1.000
Enthalpy, MJ/hr	-1.281E+06	-1.244E+06	-1.044E+05	-1.022E+05	-1.244E+06	2.435E+02	-8.153E+05	-8.137E+05
Enthalpy, mmBtu/hr	-1.214E+03	-1.179E+03	-9.898E+01	-9.688E+01	-1.179E+03	2.308E-01	-7.728E+02	-7.712E+02